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1949

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TRANSACTIONS

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Volume 183

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MINERALS BENEFICIATION
1949

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FOREWORD

With this volume of the AIME Transactions, No. 183, publication in the present size ceases. Hereafter, all volumes will be 8½ by 11½ inches, as bound, and will consist of the Transactions sections of the respective Branch journals for the preceding year. The date appearing on the volume will be that of first publication of the material, rather than the date the volume was scheduled for publication, as in the past.

Heretofore, Transactions volumes have been issued as the material in one particular field accumulates—in this case the field of mineral beneficiation. Hereafter, all of the material in the field of one of the three Branches of the Institute will be bound together. This means, in the case of mineral beneficiation, that all papers on geophysics, geology, metal and coal mining and beneficiation, fuel technology, and the nonmetallic minerals will be grouped in one volume.

Methods of concentrating ores have been improved considerably since the last volume on this subject was issued in 1946—the Richards Memorial volume on milling and concentration, No. 169. Sink-float separation is the newest of the concentration processes to be applied on a wide scale, though the principle is old, and the present volume contains three papers on this subject. Electrostatic separation has been utilized for certain special purposes for many years, but the literature on the subject is comparatively scanty. The two papers on the subject in this volume show some interesting applications of this process.

Though flotation has dominated the field among concentration processes for the better part of the last half century, reports of progress in this art continue to be more numerous than those on any other phase of concentration, and ten papers on this subject may be found in this volume. Crushing and grinding likewise afford a never-ending field for experimentation, for the subject seems to defy standardization. What holds true for one ore may well not be applicable to another even in a neighboring mine. But still, one can learn much from experience, and five groups of authors have something to say on comminution in this volume.

No volume of this subject would be complete without a few flowsheets of concentrating plants, with accompanying descriptions of the equipment used, for development of some new synthetic material may mean a radical change in a conveying system, a new alloy may enable an obsolete type of crusher again to be popular, and electronic controls at the proper places may add greatly to the smooth running and efficiency of a mill.

So here, for the thoughtful reader, is a wealth of information on the latest practice in ore concentration. May he who gains from perusing these pages be impelled, in turn, to tell others of what he himself has learned about how the balls roll around and the froth bubbles over.

E. H. Robie, *Secretary*

New York, N. Y.
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Symposium on Milling Devices and Practices

By J. F. Myers* and R. J. Tower,* Members AIME

(New York Meeting, March 1947)

"There is nothing new under the sun." All over the world, mineral-dressing engineers are working at their problems, no two of which are alike. Each encounters equipment and process problems. Many devise simple ways and means of overcoming their difficulties. Many of these problems are quite common in the industry, yet, since no one human being seems capable of thinking of everything, these problems at some plants often are looked upon as necessary evils.

The purpose of this symposium is to gather together, from competent engineers, devices and practices that have been useful to them. Most of the ideas contributed are not new, yet there is no plant operator, old timer or youngster, who will not find some idea herein that will give him a suggestion for improving his operation.

Space does not permit a complete elaboration on all of the contributions offered. Each item is summarized, but this will be ample for those skilled in the art of mineral dressing.

A list of contributors appears at the end of the paper, and the superior number at the end of each item identifies its source.

General Equipment

A signal light to show that a steel bin, containing relatively fine, moist ore, is full, can be made by suspending a copper rod in the bin to a sufficient depth to ensure ample contact with the ore. One wire of a 32-volt circuit should be attached to the rod and one to the side of the bin. Sufficient current will flow to make the signal light burn. A 110-volt circuit could be used, but 32 volts ensures safety.[10]

Fig 1 shows a circular bin for crushed ore. It is a concrete structure footed on the ground, using no supporting columns. Tailing from the sink-float was used for aggregate. Cross conveyors for discharging ore are not required, nor is a tripper for filling. Segregation is minimized by drawing from different gates. An exhaust fan on top of the bin eliminates condensation due to temperature of the ore.[21]

A beam with an adjustable point of suspension near the center permits rigging rolls for removal in a very few minutes. This makes unnecessary the lacing of the cable through the hub bolts on one side, around the shaft on the other, and the trouble of balancing the load.[8]

The ratchet handle on a Symons crusher may be replaced by a drum with a cable wound on it. When the cable is pulled by a crane or hoist it causes the drum to revolve and thus screws or unscrews the bowl. This reduces handwork and materially shortens the time required for the job.[3]

To spread ore feed from an apron feeder over a belt conveyor, place a heavy steel bar about midway between the top

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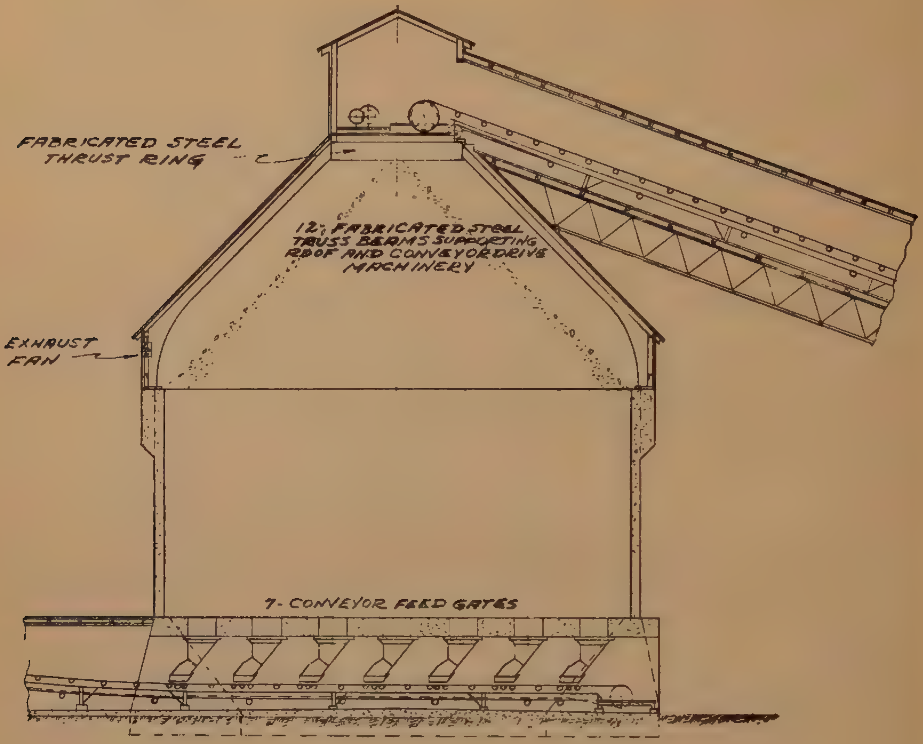
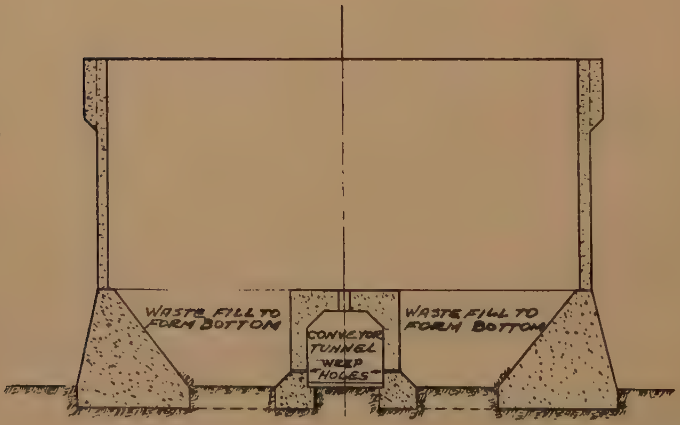
* Mill Superintendent and Assistant Chief Engineer, respectively, Tennessee Copper Co., Copperhill, Tennessee.


FABRICATED STEEL THRUST RING
12- FABRICATED STEEL TRUSS BEAMS SUPPORTING ROOF AND CONVEYOR DRIVE MACHINERY
EXHAUST FAN
7- CONVEYOR FEED GATES

SECTIONAL ELEVATION PARALLEL TO CONVEYORS



SECTIONAL ELEVATION RIGHT ANGLE TO CONVEYORS

Fig 1—Circular ore bin.

and bottom of the ore ribbon being discharged from the feeder. This nullifies the breaking off of the feeder load and eliminates surges.[1]

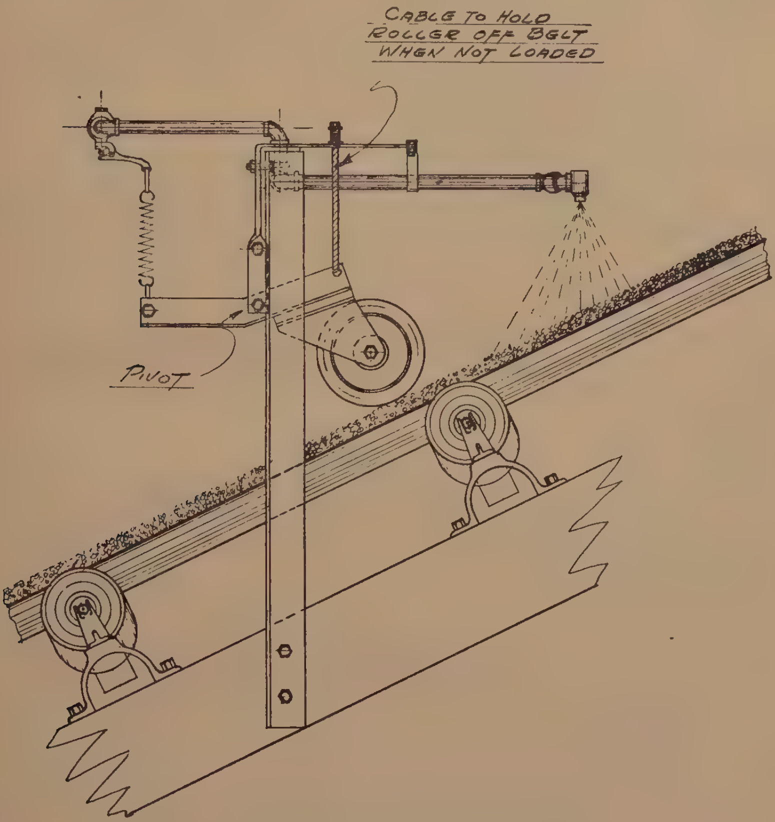


FIG 2—AUTOMATIC CUTOFF FOR BELT SPRAY.

Fig 2 shows an automatic means of cutting off sprays on a conveyor belt when the belt is not loaded. This prevents formation of pools of water around the tail pulley of inclined belts.[20]

A good way to clean a belt handling sticky ore is to place a water spray followed by a spring, or weighted wiper of solid rubber, on the return side of the conveyor between the head and snubber pulleys.[1]

A ¼-in. wedge placed under the rear foot of troughing idlers, to tilt them in the direction of travel, will improve alignment of conveyor belts.[1]

Inclined conveyors handling coarse rock, fed with a step-bottom chute with V-bottom to put the load in the center of the belt, must have tight fitting skirts. Otherwise large pieces may roll back under the chute at the center of the V and jam between the chute and the edges of the belt, where there is less clearance.[4]

A very small amount of commercial castor oil added to the inner side of a conveyor belt gradually builds up an ideal traction surface. It prevents the belt from slipping and prolongs its life. If any part of the carcass is exposed, the oil penetrates the fabric and aids in preserving it. Too much oil will cause the belt to slip, and it must not be used on a stalled belt nor on one that is already slipping.[10]

To absorb shock and prevent damage to a conveyor belt caused by large pieces of ore falling on it, place under the frame a pneumatic cushion 1½ in. thick and the length and width of the carrier frame. Inflate cushion to suitable pressure. Provide vertical guides for carrier frame, and anchor it against belt drag with flexible arms.[5]

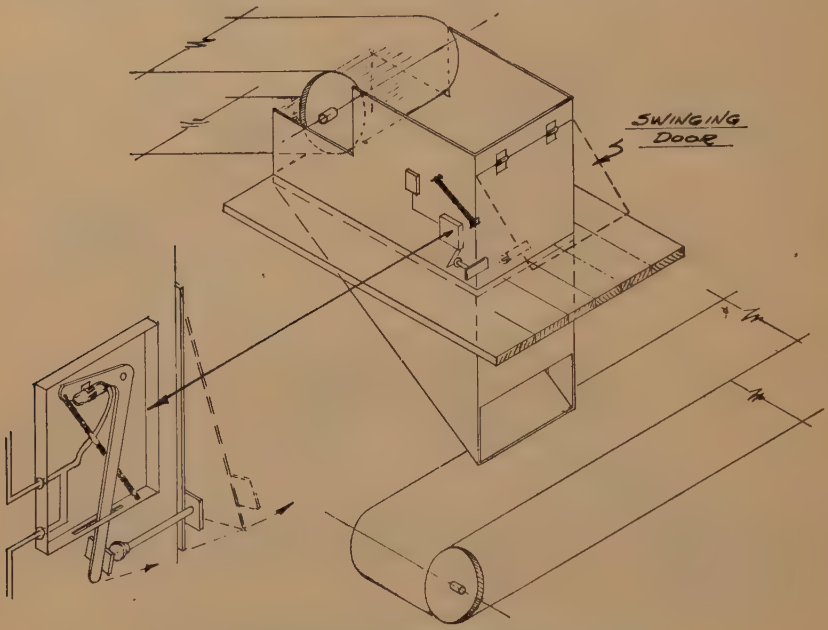


FIG 3—DEVICE TO STOP CONVEYOR IF CHUTE DIKES.

Another means of preventing injury to a conveyor belt is to place a grizzly in the transfer chute, so that the fines reach the belt first and form a protective covering for it. If the ore is too sticky to pass the grizzly, hang a chain curtain in the chute, to reduce the velocity of falling chunks.[1]

To eliminate expensive clean-up and damage to conveyor belts caused by diked or plugged transfer chutes, the dust hood of the chute can be equipped with a hinged gate *A* as in Fig 3. When the chute becomes diked, the weight of material against the hinged gate causes it to open and trip mercoid switch at *B*, which stops the conveyor. By interlocking conveyors, feeders, crushers, and other machinery feeding the conveyor, all the system involved will be stopped automatically.[15]

Durable liners for ball mills can be made by welding ½ in. by 3½-in. steel bars, set on edge about 1½ in. apart, to a ⅝-in. plate the size of the liner. The space between the bars will fill up with ball fragments, to take the wear. Size and spacing of bars depend on size of balls used.[2]

To maintain grinding capacity, which is reduced when new liners are installed in a grinding mill, because of the reduction in diameter and peripheral speed, a larger motor pulley may be used on a belt-driven mill. Changing the pulley is facilitated by using a demountable rim type with a permanent hub. Motor take-up must be provided to accommodate change in pulley centers.[7]

A drop-bottom bucket can be used to charge a mill. The bucket is handled by an overhead crane and is so arranged that the bottom opens when the bucket is set down on the mill door.[3]

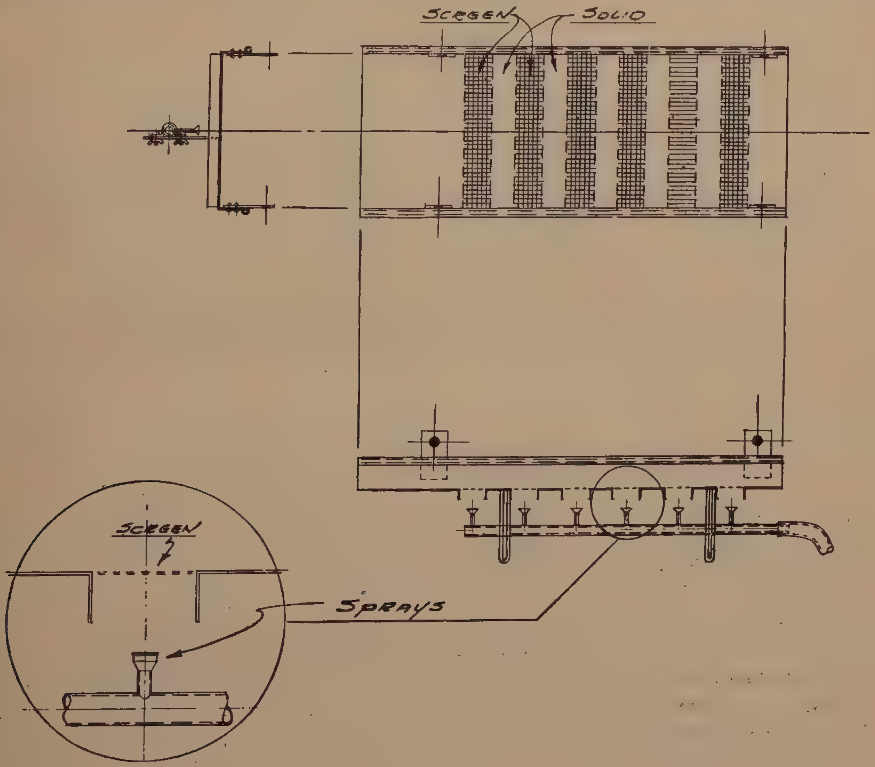


FIG 4—LAUNDER SCREEN.

Welding blades to the channels of classifier rakes causes the channels to bow, but counterbowing them before welding will result in straight channels after welding.[3]

Because of ineffectiveness of screening with screen cloth having openings smaller than 48-mesh, and the difficulties of sizing by hydraulic classification when pulp contains minerals ranging from 2.65 to 7.1 sp gr, a launder-type stationary screen has been developed, which is quite satisfactory. The unit (Fig 4) consists of a launder whose deck is made up of alternate sections of screen cloth and solid areas. Sprays are placed below the screen-cloth sections, and water is played on the screens momentarily at intervals. Drip plates are attached to the underside of the screen sections at the lower ends, to prevent the undersize from falling on the spray nozzles and choking them. Such a unit is reported to have recovered in the oversize 97 pct of the plus 100-mesh material, 84 pct of plus 150-mesh, and 75 pct of the minus 200-mesh, whereas 65 pct of the total minus 325-mesh was eliminated in the undersize.[17]

A rope coupling can be used for a mill or crusher drive, and it will protect them against shock. In a crusher it will prevent breakage caused by the presence of tramp steel in the crusher feed. Coupling is made up of two halves, each half consisting of a face plate into which pins have been pressed. The motor half is larger in diameter, and the pins in the driven half are set on a diameter small enough not to touch those on the motor half, when the halves are put together. The pins are grooved, and an endless manila rope belt is laced alternately around the pins of the two halves. Fig 5 shows the details.[17]

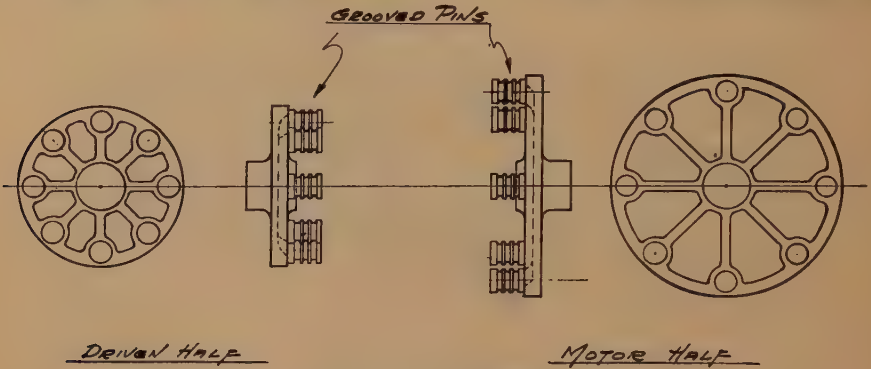


Fig 5—Rope coupling.

A small spitzkasten-type hydraulic classifier, comprising a settling box with an intermittent sand discharge, is operated in closed circuit with fine-crushing rod mills, to return a small amount of the coarser sand and tramp oversize to the rod-mill feed. This practice eliminates troublesome accumulation of oversize material in the fine-crushing circuit.[18]

Where large volumes of tough, voluminous froth are produced, launders are not used. The flotation machines are set over bins that have sloping bottoms and control gates, to direct the products to the desired places. Properly applied sprays help to control the froth.[11]

Usually the maintenance of the proper level of the pulp in a pneumatic flotation machine is done by manual operation of the tail gate. A device has been developed, which, after initial setting, does this automatically. The level-controlling device is shown in Fig 6 and consists of a motor *A*, speed reducer *B*, tail gate assembly *C*, swinging gate *D* at the discharge end of the concentrate launder, and adjustable mercoid switches *E*. Switchboard *F* contains time relays, reversing switch, and switches for manual and automatic operation. The time relays control the frequency travel during the interval.

The controlling principle of the device is that a change in pulp level provides a corresponding variation in volume of froth overflow. The froth stream, impinging on the swinging gate at the end of the concentrate launder, changes the deflection of the gate and thereby changes the position of the mercoid switches. Deviation of these switches from their neutral position selects the direction in which the driving motor will rotate during the next operating interval of the timer. Thus the gate is raised or lowered gradually until the volume of the froth overflow assumes its normal proportion, and the mercoids return to a neutral position.

To make the initial setting, adjust the pulp level manually, and set mercoid switches in neutral position by means of adjustable lever arm *G*. If the controller now consistently tends to drop the froth level, it indicates that a decrease in the frothing agent is needed. Increase in frother would be indicated by tendency of controller to maintain high pulp levels.[15]

In flotation machines of the Forrester,

Southwestern and Munro-Pearse types, it is necessary to use sufficient air to keep all the particles of ore in suspension. Otherwise these machines sand up after a few hours of operation. If less air is used, a small amount of tramp ore or oversize from the classifier crawls along the bottom of the machine and gradually begins to build up from the tailing end. An airlift installed in the cell at the tail gate will remove this material and permit an appreciable saving of air. The airlift is operated with the same air that is used in the machine. This is far more positive than sand holes, which are sometimes used. Where coarse feed is used in mechanical cells, such an airlift is practical in each cell.[10]

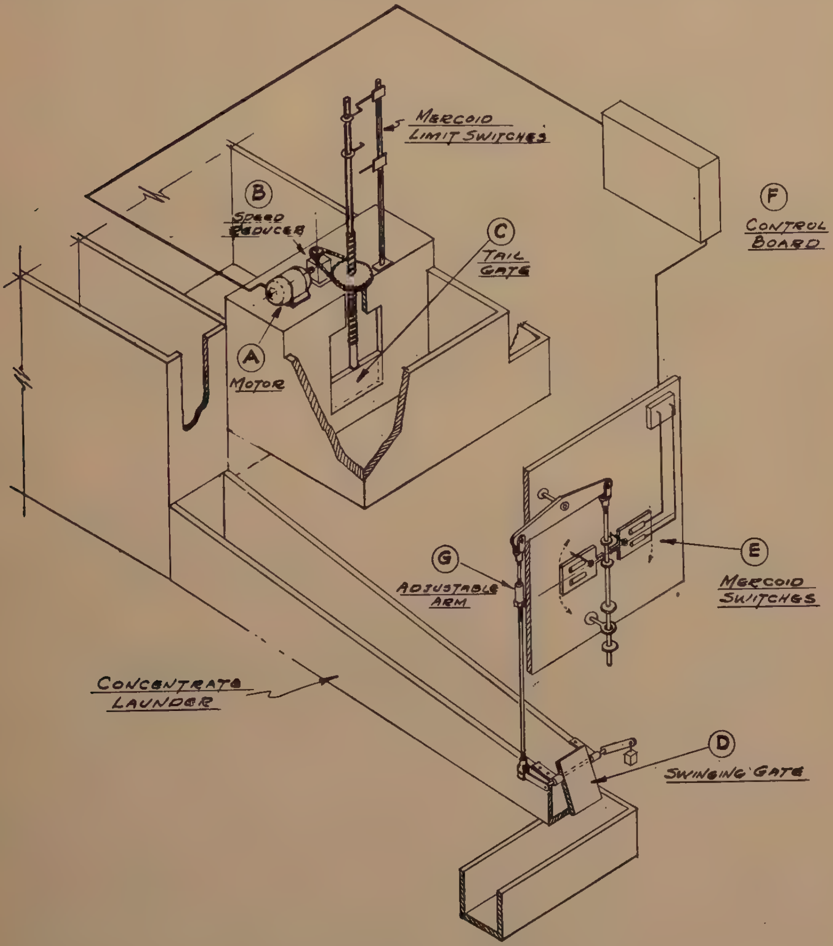


FIG 6—PULP-LEVEL CONTROLLER.

On deep, 10-ft pneumatic cells of the Munro-Pearse type, power failure may cause sanded up conditions that prevent the cells from coming back into operation when power is resumed. Hydraulic jacks to lift the air header and aerating pipes out of the cell, and to lower them back slowly when power comes on again, make it possible for the cells to be operating within a few minutes. Electric, four-way cocks, interlocked with the blower motor, make the operation complete.[10]

A suitable plug (Fig 7) for quick, easy draining of flotation machines consists of a wedge-shaped wooden block whose face is covered with rubber launder lining. The plug slides in metal guides and has a handle. It has none of the disadvantages, such as seizing, swelling, and sudden release, common to various types of plugs, flap valves, and plugcocks. It is cheap and easy to make. It is easy placed or removed, and it can be made for any size of opening.[17]

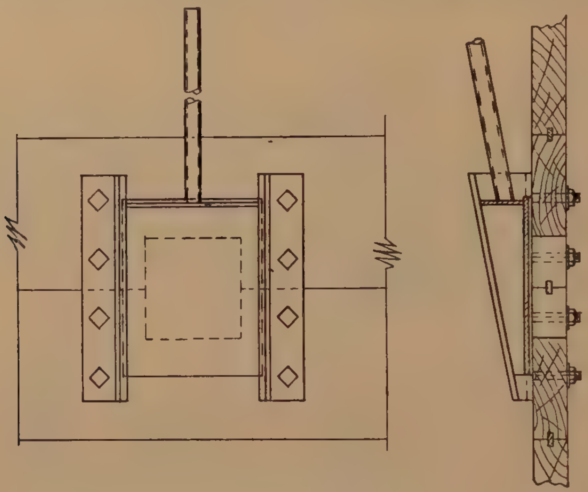


Fig 7—Plug valve.

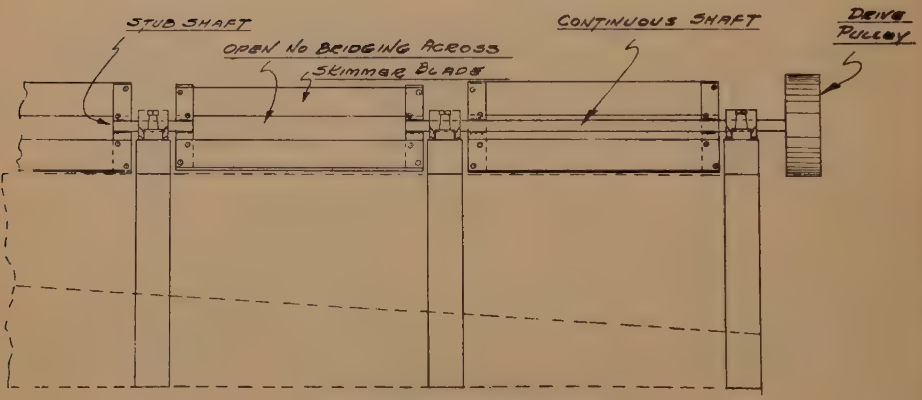


Fig 8—Stub skimmer shaft.

Instead of continuous froth-skimmer shafts running the entire length of the machines, short stub shafts can be welded to the skimmer-blade arms as shown in Fig 8. This prevents concentrate from bridging over between opposite blades and allowing large blobs to fall back into the machine. Clamps for fastening the skimmer blades to the continuous shaft are unnecessary, and there is considerable saving in material. At the ends of the machine the skimmer shaft, carrying the drive pulley, extends the full width of the last cell.[17]

To check possible metallurgical results from heavy-media process of ores of plus ¼-in., stir up a mixture of minus 100-mesh ferrosilicon and water. Weigh a known volume of the mixture and add ferrosilicon or water to obtain the desired specific gravity. Drop a screened sample of the ore into the mixture, stir rapidly and skim off the float, which is the tailing. Wash both the sink and float on a screen, weigh and assay.[1]

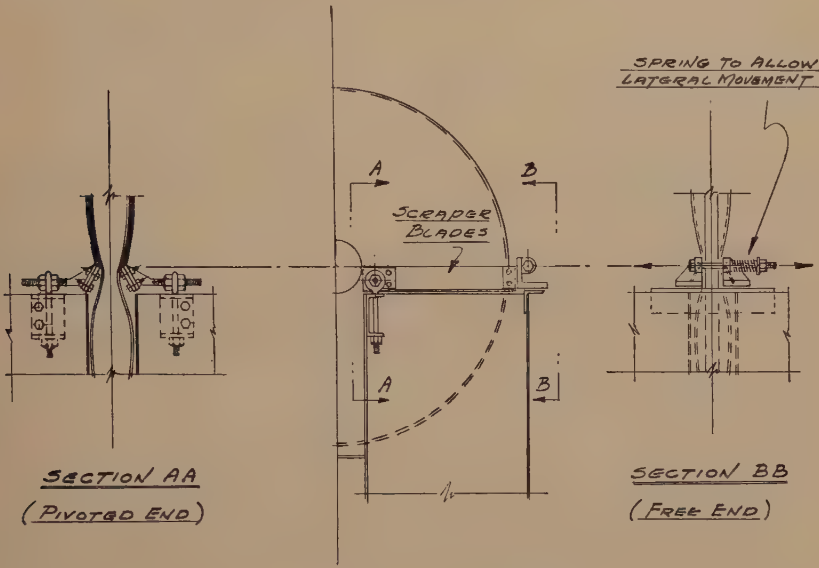


FIG 9—SCRAPER FOR DISK FILTER.

In the heavy-media process a meter can be used to record the differential in pressure between two tubes of different lengths immersed in the ferrosilicon medium. A recording 24-hr chart is calibrated to indicate this differential in terms of specific gravity. This eliminates the taking of samples by hand, and corresponding chance for error in the record.[13]

On vertical leaf filters, for instance the American, it is quite difficult to keep the various segments lined up, particularly after the filter has been in service for a long time. This causes uneven pressure on the filter cloth of the segments by the scrapers as the filter cake is discharged. Scrapers usually are bolted to the frame. Fig 9 shows a means of mounting the scrapers, so that they are free to move sideways as they follow the movement of the out-of-line filter segments, and thus increase the life of the filter cloth.[10]

AUXILIARY APPARATUS

When pumping is severe, pumps can be equipped with ammeters, and when the current falls off to a certain point, it will indicate that parts are worn and should be replaced. This leads to low power and replacement costs and eliminates pump failures.[3]

Worn sand-pump casings can be repaired with alundum or carborundum-cement mixtures. After it has cured for 90 days, a repaired case will give from one to two times the service of a new case, and the cost of repairing is about one fifth that of a new one.[3]

To stop pump surges, insert a short piece of small pipe with a valve on the end into the suction of the pump, and admit the amount of air found necessary.[12]

A pump box that cannot be seen by the operator can be equipped with an electronic high-level control and an alarm, to warn the operator of a rising pulp level.[4]

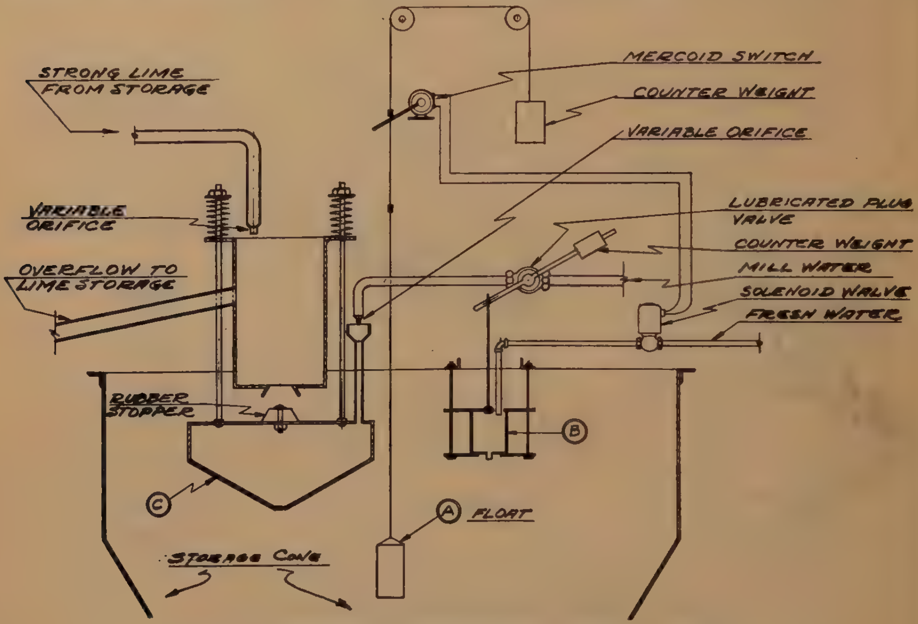


FIG 10—DEVICE FOR DILUTING LIME FOR UNIFORM STRENGTH.

Worn or broken rubber diaphragms for pumps can be repaired in the same way in which conveyor belts are patched and vulcanized. Repaired diaphragms give an extra life of about 30 pct of the original and may be repaired several times.[9]

When less air delivery is required, throttling the suction side of a centrifugal blower will save power. Throttling the discharge side does not.[8]

A heavy cylinder-stock oil with wool waste is often used for lubricating trunnion bearings of grinding mills. A little light engine oil fed automatically by a drip feeder will reduce the power consumption 5 pct or more.[10]

Lubricating grease in flotation pulp causes froth trouble, while lubricating oil is much less injurious. Bottle oilers that are automatic, and feed oil only when shaft is in motion, should be used on classifiers, diaphragm pumps, tank drives, etc., to reduce this trouble.[10]

Oil drainage from crushers, pumps, gear cases, and other equipment may be treated for re-use by adding 1 lb caustic soda to 40 gal of oil, boiling for 15 minutes, and allowed to stand overnight. Wastage will have settled in the bottom of the container.[2]

A miniature sample cutter that re-cuts the stream from the main cutter reduces the bulk of the sample, thereby saving considerable work.[9]

A device used to furnish a supply of lime and water of uniform strength is shown in Fig 10. Operation is as follows: As the level falls in the storage cone, float *A* actuates the mercoid switch, which energizes a solenoid valve and allows fresh water to enter container *B* at a rate greater than the discharge from it. The weight of water in container *B* opens a plug cock in the mill-water line and allows mill water to enter container *C* at a rate greater than that of the discharge from it. The weight of water in container *C* overcomes the effect of the suspension spring and disengages the rubber plug in the bottom of the strong-lime vessel. Thus strong lime is discharged into the storage cone and mixed with dilution water. Quantities of lime and mill water are regulated by the size of the orifices. When the level in the storage cone rises sufficiently, the sequence is reversed and the flow into the cone stops.[20]

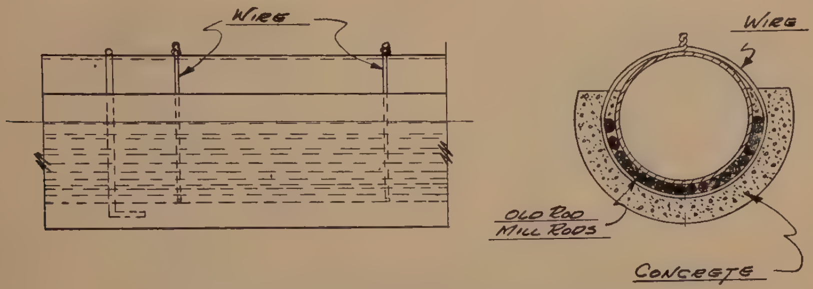


Fig 11—Method of repairing pipes carrying pulp.

A method of repairing pipe lines carrying flotation pulps, while the lines are in service, is shown in Fig 11. Fittings are repaired in the same manner. Costs are about equal to new.[14]

A laboratory screw-type classifier can be made by using a conventional screw conveyor in a steel trough with standard head and tail ends. All parts may be purchased from any conveyor manufacturer, and drive and water boxes provided as desired.[1]

A downdraft hood is quite satisfactory for the control of dust in sample-preparation rooms. Downdraft takes advantage of the natural tendency of dust to settle and avoids drawing it past the workman's face, as is done with the overhead type of hood.[21]

The recommended procedure for infrasizing has been to charge the sample into the first cone (smallest diameter) and to operate at a constant air velocity of 20 in. on the water manometer for a standard time period. A modified procedure has been used over the past two years, which has several advantages over that recommended by Haultain. The sample is first charged into cone No. 6 and elutriated until the rate of increase in weight of bag product becomes small. This determination is aided by placing a short length of glass tubing in the line leading to the bag. Two or three weighings of the bag can be made when the amount of solids entering it diminishes. A convenient standard is an increment of less than 0.5 pct in 15 minutes. When the desliming operation is completed, the sample is removed from cone 6, charged into cone 1, and sized as usual.

This procedure permits better scrubbing in cone 6 for release of fines. It eliminates coating of the other cones by fines. In addition, it is possible to reduce the air velocity and obtain still finer bag products; as, for example, at velocities of 5 and 10 in. The sizing is started at 5 in.; when this is completed the bag is changed and the velocity increased to 10 in; on completion the bag is again changed, and desliming continues at 20 in.; following this, sizing is completed from cone 1. While it is possible to operate at these lower velocities with the older procedure, the lack of effective scrubbing in cone No. 6 prolongs the separation needlessly.

With normal mill products the extra infrasizer fractions may be of no importance, but where the minus 20-in. bag product is a large fraction of the original material, and contains a significant amount of valuable material, its further fractionation may be important.[22]

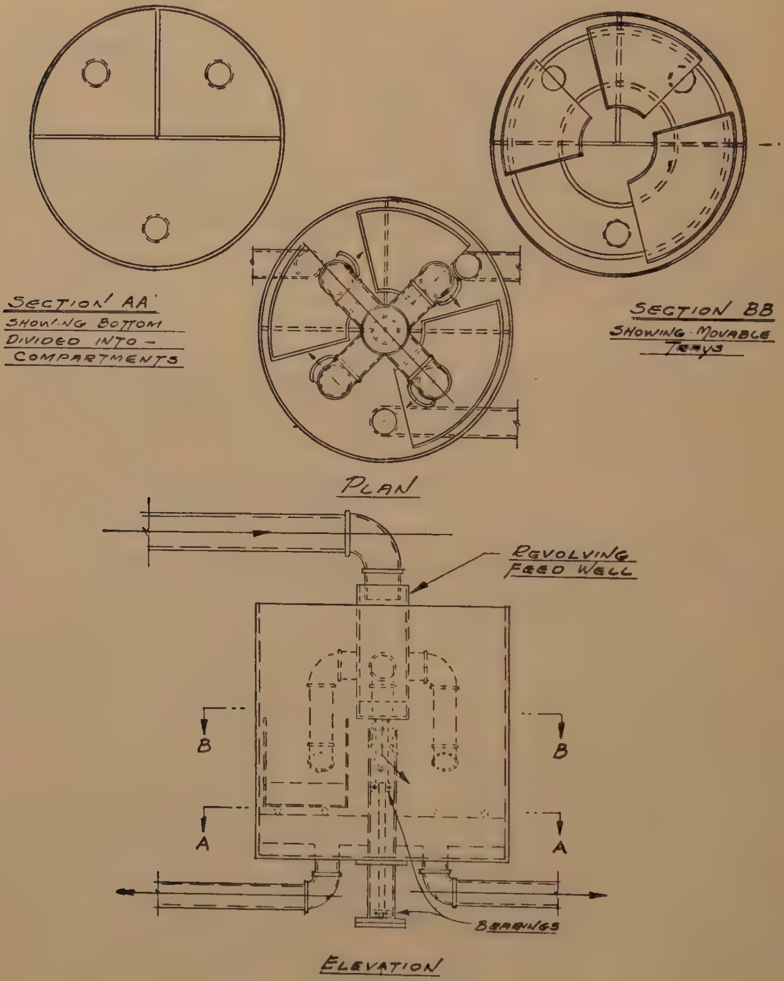


Fig 12—Adjustable distributor box.

Although satisfactory laboratory flotation cells of the agitation and subaeration type are available, the same can hardly be said of pneumatic machines. In part this is because of unwitting acceptance of an unsound principle of model construction; namely, that of reducing the device to be imitated equally (or nearly so) in all dimensions. It would be preferable to adopt for model building the principle of reducing dimensions that are relatively unimportant and keeping others at nearly full scale. In keeping with this sounder principle, a glass pneumatic cell has been constructed in which the depth of pulp is nearly 3 ft, the length about 3 ft and the width 1 or 2 in. Circulation of pulp and froth is obtained by the use of longitudinal partitions made of plate glass. The cell has the shape of a truncated inverted wedge, so that it is somewhat wider at the top than at the bottom.

Control of the level of the pulp body is obtained by a gooseneck device that can be raised or lowered by a screw thread of moderate pitch. Control of the width of the bubble column is obtained by a glass wedge dipping in the bubble column and capable of vertical motion through a screw thread of moderate pitch. Taken together, these two control devices permit the regulation, at will, of the relative volumes of pulp body to bubble column, as well as the shape of the bubble column. The glass and plastic construction minimizes the effect of wall-derived salts, and the retention of depth and length at the expense of width permits the bubble to imitate better in length of passage through the pulp and in pressure changes the behavior of bubbles in industrial pneumatic cells.[6]

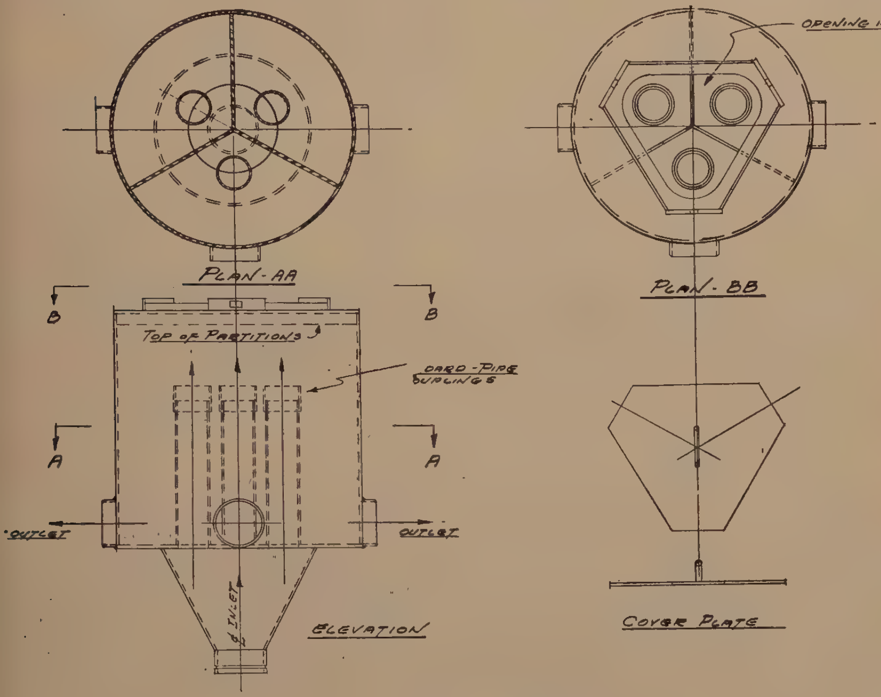


FIG 13—PULP DISTRIBUTOR.

Moving pictures have been used to study comminution, the operation being carried out in a short ball mill fitted with a coarse screen end, through which the pictures were taken. In the present device stroboscopic viewing replaces moving-picture photography, and plate-glass ends replace the screen ends used heretofore. The mill is 18 in. inside diameter and 3 in. inside length, and is mounted on steel rollers driven by a variable-speed motor. The stroboscopic illumination is obtained by the use of a strobotac with strobolux attachment. This device has shown that there is differential movement of different layers of balls, and that much grinding must occur within the ball charge. For the study of wet grinding, an isotropic mineral is preferred, with a liquid having the same index of refraction. This makes visible the action of the suspension, and obscures ball motion much less than any other medium.[6]

A pipe nipple may be fastened to a rubber hose without the use of clamps, by heating the nipple and forcing it into the end of the hose. The heat vulcanizes the hose to the nipple.[8]

To secure anchor bolts to old concrete foundations, drill holes in the concrete, insert bolts, and fill holes with molten sulphur.[1]

A distributor box with adjustable discharge is shown in Fig 12. The bottom of the distributor is divided into three compartments, each with one discharge pipe. Three movable trays are supported above these compartments. The trays have one closed and one open end. If more pulp is desired in one compartment, the closed end of the movable tray for that compartment is moved to cover part of the next compartment. The feed well is caused to revolve by the action of the pulp discharging from the 45° elbows. One application of this device is to vary the feed to grinding mills, as their capacity changes because of different degrees of liner wear.[10]

Another type of pulp distributor is shown in Fig 13. This machine has no moving parts and gives an equal distribution to each of its component partitions.[16]

Often it becomes necessary to blast large pieces of ore in jaw crushers, feed chutes and ore bins. This creates a dust that is undesirable from the point of view of health and clean up. To allay this dust and control its ill effects, a dense fog of water can be generated before the blast by a "bazooka," illustrated in Fig 14. This device is made of standard pipe fittings. Several manufacturers make devices for the same purpose in more elaborate forms.[16]

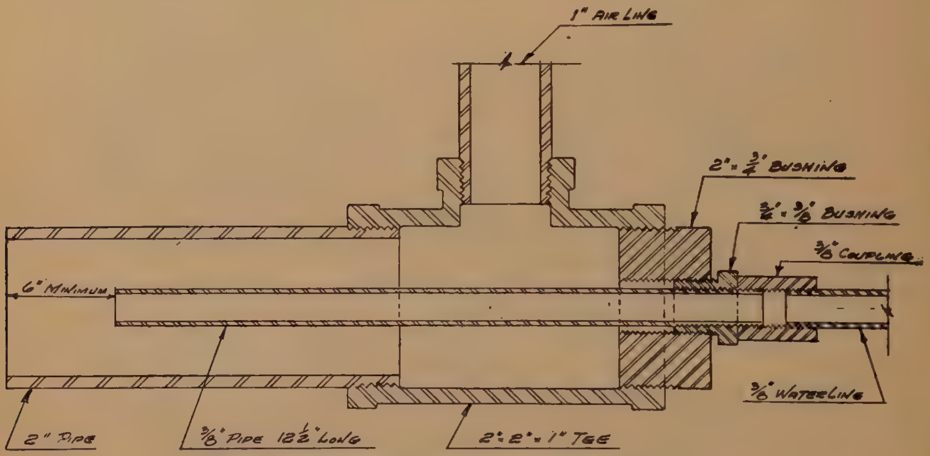


Fig 14—Air and water blast (2-inch bazooka).

Concentration Practices

Sometimes where a two-stage grinding circuit is used, it is possible to reduce the tonnage lost on account of mill or classifier repairs. Thus, if a primary classifier is shut down, the tonnage can be reduced by a fixed amount, and the same grinding maintained on the secondary circuit. If one primary mill is temporarily removed from service, a somewhat reduced tonnage can be handled by increasing the tonnage on the adjacent unit while still maintaining the same grind on the two secondaries. If a secondary unit is down for repairs, a reduced tonnage is treated by the two primaries and one secondary. Switches of this kind require special launder or splitter-box arrangements, but over a period of time it is surprising how much extra tonnage it is possible to handle. A few trial runs will establish the tonnage that can be accommodated by the different combinations.[9]

Fine grinding and recleaned concentrate sometimes make a product in which the silica content must be regulated. This can be done by pumping the desired amount of siliceous sand into the concentrate from a pocket in the tail line.[9]

Breakdowns can be kept to a minimum by using a system of inspection, covering every possible detail of maintenance. Necessary repairs are made on planned shutdowns of portions of the operation based on these inspections. Such a system has resulted in operating time averaging 95 pct.[14]

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DISCUSSION

(*E. H. Rose presiding*)

Harlowe Hardinge*—Mr. E. H. Rose, in asking me about our "electric ear," sound control for ball mills, knows that I am one who is very much in favor of controls in general, and ways and means of eliminating the personal equation when it comes to operating grinding mills and other devices. I made a trip throughout the United States recently, and in talking to operators in various industries, I was impressed by the amount of investigation being made by them on ways and means to simplify their operations, eliminate unskilled labor and increase their overall operating efficiency.

Several years ago, control features were classed as gadgets to "get out of order;" but today, there seems to be a distinct change in sentiment, perhaps because some of these controls actually do work and pay dividends to the user. It is important, however, that the control be simple and effective within the normal limits of operation. So far as possible, a control should start at the top of the operation rather than at the bottom. In other words, "it is much easier to prevent the disease than to cure it."

Those controls that regulate the material or operation at the source have a much better chance of continuing to function than those that try to correct an unbalanced condition long after this trouble has started. It seems to me that this, fundamentally, is the method of approach that should be followed so far as possible. When we are dealing with nature and nature's variables, it is impossible to achieve perfection. Controls, at best, are compromises.

While the present paper brings out ideas and methods which are well known to some it will suggest to the operator that there may be other ways of getting over certain difficulties.

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Milling Lead-zinc Ores at Iron King Mine, Prescott, Arizona

By H. R. Hendricks,* Member AIME

The ore of the Iron King mine, being very hard and having a very fine crystalline structure, presents many problems in milling that are not present in ordinary lead-zinc ores. This very fine crystallization causes much interlocking of minerals and is most prevalent between the gold and pyrite and the sphalerite and pyrite.

The recovery of gold is extremely important in the economic value of the ore and therefore constitutes a large part of the metallurgical problem to be solved. All of the gold is free but it has not been observed in any of the mill products except under very high magnification. Some of the gold is tarnished, and this condition also makes recovery difficult. The gold values have been determined to be highest in association with chalcopyrite and arsenopyrite.

The silver occurs mostly in association with galena and tennantite. A small amount of native silver has also been observed occasionally. The galena, being softer than the other minerals, frees much better in grinding. Some oxidation of galena is present in the ore, as the nonsulphide lead content amounts to approximately 10 pct of the total lead. Some oxidation of sphalerite is also present in the ore.

Microscopic examination of the zinc concentrate shows many grains in the minus 325-mesh to be half pyrite, together with the fact that some of the zinc is marmatite, makes production of a high-grade zinc concentrate very difficult unless recovery is sacrificed.

In the grinding, it has been determined that the flotation feed must be maintained at 85 pct minus 200-mesh to free the minerals sufficiently to make the best economic recoveries. Table 1 illustrates the effect of fine grinding. It shows that the bulk of the recovered minerals is in the minus 325-mesh size and is the part of the minerals that is more readily freed. The pyrite concentrate is not given in the table because it contains practically no minus 325-mesh.

Development of Present Flowsheet and Milling Practice

The first milling operation started in 1938, with a mill capacity of 140 tons daily, producing a bulk gold-silver-lead concentrate. After several stages of expansion and changes of the mill flowsheet, the mill is currently treating 380 tons of ore daily and three separate concentrates are produced. In order of their importance and value, these are: (1) lead concentrate, (2) zinc concentrate and (3) pyrite concentrate.

The present average grade of mill heads is: gold 0.11 oz, silver 4.00 oz, lead 2.50 pct, zinc 7.60 pct, copper 0.22 pct, iron 22 pct and 30 pct insolubles. Specific gravity of the ore is 3.8.

During the period when bulk flotation was practiced, the flowsheet was very simple. The ore was ground fine enough to liberate as much gold, silver and lead as possible. Necessary collectors and frothers were added in sufficient quantity

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to remove the most active minerals. This produced a concentrate that would contain most of the gold, silver and lead. On account of the fine interlocking between the gold and pyrite, the recovery of gold by flotation was not entirely satisfactory, because the pyrite was approximately half of the mill feed and if all the pyrite were recovered the grade of concentrates would be too low. Laboratory research then indicated that differential flotation would produce a zinc concentrate that could be marketed at a slight profit. The mill was then converted, in the latter part of 1939, to differential flotation, and the capacity simultaneously was increased to 225 tons daily.

TABLE 1—*Distribution of Metals in Mill Products*

PER CENT

Mesh	Wt.	Au	Ag	Pb	Zn	Cu	Fe	Insol.
HEADS								
+100	1.88	0.9	1.4	1.1	0.9	1.0	0.5	3.6
+200	13.23	8.5	8.2	5.7	7.1	7.0	5.5	21.7
+325	20.10	24.3	15.6	11.9	17.8	16.1	22.3	16.5
−325	64.79	66.3	74.8	81.3	74.2	75.9	71.7	58.2
Total	100.00	100.0	100.0	100.0	100.0	100.0	100.0	100.0
LEAD CONCENTRATES								
+200	1.00	1.0	0.8	0.4	1.0	0.8	1.0	4.0
+325	12.38	13.3	9.0	6.6	14.4	11.5	15.4	27.4
−325	86.62	85.7	90.2	93.0	84.6	87.7	83.6	68.6
Total	100.00	100.0	100.0	100.0	100.0	100.0	100.0	100.0
ZINC CONCENTRATES								
+200	1.23	1.0	0.7	0.6	1.0	1.1	1.1	2.9
+325	11.34	11.2	7.4	5.6	10.3	13.5	14.3	23.4
−325	87.43	87.8	91.9	93.8	88.7	75.4	84.6	73.7
Total	100.00	100.0	100.0	100.0	100.0	100.0	100.0	100.0
TAILS								
+100	2.0	1.0	1.2	2.0	1.3	2.2	0.7	3.8
+200	12.8	7.9	14.6	12.7	15.6	11.1	6.1	20.0
+325	13.9	16.9	14.8	15.3	23.8	15.6	14.7	12.4
−325	71.3	74.2	69.4	70.0	59.3	71.1	78.5	63.8
Total	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0

A modern cyanide plant was added to the circuit in the latter part of 1940 to treat the zinc tailings for additional recovery of gold, and was operated continuously until the latter part of 1942. At that time a market developed for pyrite concentrates that would make a suitable smelter flux. Since there was still much gold locked with pyrite, it was found that a pyrite concentrate could be produced that would have a slightly higher economic value than the product of the cyanide operation. The latter was then discontinued, after the installation of Wilfley tables and production of the pyrite concentrate began. This pyrite concentrate was first produced by tabling the zinc tailings. Later it was found that tabling the lead tailings before zinc flotation would benefit the zinc circuit, because the removal of some of the pyrite would reduce the circulating load.

Again in 1943 the mill capacity was increased, by the addition of another ball mill, classifier and additional flotation cells. The Wilfley tables producing the pyrite concentrate were double-decked to handle the additional tonnage and thereby produce a larger amount of the pyrite concentrate.

PRESENT FLOWSHEET (FIG 1)

The ore as received from the mine is placed in a 100-ton bin near the shaft. From this bin the ore is transferred by a 30-in. apron feeder to a 24-in. conveyor belt, which delivers the mine-run ore over a grizzly set at 1½ in. A 40-ton bin for receiving custom ore from trucks also feeds onto the grizzly. The grizzly undersize drops onto an 18-in. conveyor belt just behind the primary crusher. The grizzly oversize passes into a 15 by 30-in. Buchanan Blake crusher, which reduces it to minus 2-in. size, then it joins the grizzly undersize on an 18-in. conveyor belt. This conveyor delivers the ore over a magnetic pulley, for removal of tramp iron, onto a vibrating screen with ¾-in. openings. The screen undersize drops onto an 18-in. conveyor belt behind the secondary crusher and the oversize is delivered to a 24-in. Symons shorthead cone crusher, where it is reduced to ¾-in. size and joins the screen undersize on the same conveyor. This conveyor carries the ore over an electric weighing scale and delivers it to a 500-ton round steel storage bin. Two 18-in. conveyor belts under the bin deliver the crushed ore to the two 75 Marcy grate-discharge ball mills.

Each Marcy ball mill is operated in closed circuit with a Dorr classifier. Reagents added to the grinding circuit are zinc sulphate, sodium cyanide and thiocarbanilide. The density of the classifier overflow pulp is maintained at 20 pct solids. The overflow from the two Dorr classifiers joins and is pumped, by a 4-in. Wilfley sand pump, to the flotation circuit on the floor above the ball mills and classifiers. This pump delivers the pulp to No. 3 cell of a 10-cell bank of Denver Sub-A No. 21 flotation machines. Cresylic acid and Z3 are the reagents added to the head of the lead flotation circuit.

The rougher lead concentrates from cells 3 to 10 inclusive are laundered to No. 1 cell for cleaning. The middling from No. 1 cell enters No. 2 cell, where some additional concentrate is removed before the product is returned to No. 3 cell to join the original feed.

The lead concentrate is laundered to a 12 by 8-ft thickener and, after thickening, is lifted by a 3-in. Denver diaphragm pump to a 4-ft, three-leaf Eimco disk filter. The filtered lead concentrate is discharged into a 40-ton wooden storage bin with chinaman chute bottom arranged for easy loading into trucks, which transport the concentrate to the railroad siding, a distance of approximately two miles, where it is loaded into gondola cars for shipment to the smelter.

The lead tailing is pumped by a 4-in. Wilfley pump to a distributor for three double-decked Wilfley No. 6 tables, where a pyrite concentrate is removed. This pyrite concentrate is pumped directly to a filter without thickening. The filtered pyrite concentrate drops into a 30-ton bin similar to that used for lead concentrates, and is also trucked to the railroad and loaded into gondola cars.

The table tailings flow by gravity to a 40-ft thickener, for recovery of water. This recovered water is returned to the grinding circuit for classifier dilution. The underflow from this thickener enters a 6 by 6-ft conditioner at approximately 45 pct solids. Hydrated lime and copper sulphate are added to the pulp in this conditioner. The overflow from this conditioner enters a second 6 by 6-ft conditioner along with the middlings from the zinc cleaner circuit. The return of the middlings from the zinc cleaning circuit dilutes the second conditioner overflow to 30 pct solids for zinc flotation. Z3 is added to the second conditioner overflow.

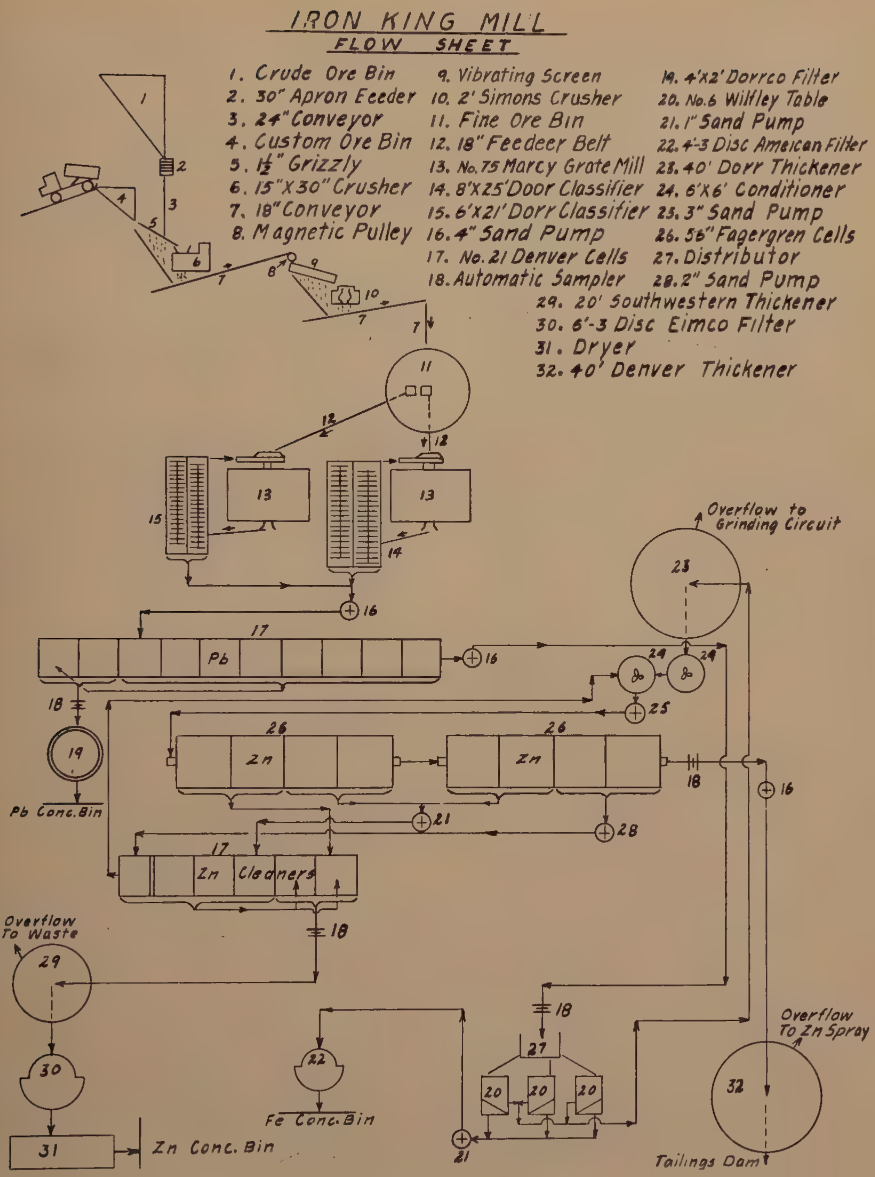

FIG 1—FLOWSHEET, IRON KING MILL.

The conditioned pulp is delivered by a 3-in. Wilfley sand pump to No. 1 cell of an eight-cell bank of Fagergren 56-in. flotation cells. Zinc flotation is accomplished in two separate circuits. The Fagergren cells are used for roughing and a six-cell bank of Denver Sub-A No. 21 flotation machines is used for cleaning.

The zinc concentrate from No. 1 and No. 2 cells of the rougher is laundered directly to No. 1 cell of the cleaners. The zinc concentrate from cells Nos. 3 to 6 (some additional Z3 being added as the pulp enters No. 5 cell) is pumped by a 2-in. Wilfley sand pump to No. 3 cell of the cleaner circuit. The concentrate from rougher cells 7 and 8 is pumped to No. 6 cell of the cleaner circuit. Concentrates from cells 3 to 6, inclusive, are laundered to Nos. 1 and 2 cells for second cleaning.

All tailings from the cleaner circuit return by gravity to the No. 2 conditioner. Cleaned concentrates are laundered to a 20-ft thickener. The thickened concentrate is lifted by a 3-in. Denver diaphragm pump to a 6-ft, three-leaf Eimco filter.

The filter cake carries 20 pct moisture because of the extremely fine grind. This moisture is reduced to 8 pct by passing the cake over a pan drier equipped with continuous drag, after which it is distributed to either of two 30-ton storage bins beneath the drier. The zinc concentrate bins are similar to the other concentrate bins and concentrates are shipped in the same manner.

Tailings from the zinc rougher circuit are pumped to a 40-ft thickener for recovery of water to be used on zinc sprays. The use of tailing water has been found beneficial to the zinc cleaning because its alkalinity is higher than that of fresh water.

The thickened tailings flow by gravity to tailings-storage dam.

Milling Practice at New Lead-zinc Concentrator of Phelps Dodge Corporation

By R. C. Thompson*

The lead-zinc mill of Phelps Dodge Corporation, Copper Queen Branch, Mines Division, Bisbee, Arizona, is about 3 miles from the main hoisting shafts of the Junction and Campbell mines at Lowell. All the ore treated in the mill is obtained from these two mines.

The mill was designed for an all-flotation treatment of 450 tons of lead-zinc ore per day, and has been operated at that capacity since the start of milling operations on Nov. 17, 1945. An expansion program is underway whereby the mill capacity will be doubled, but this article describes only the original program and presents results of the 450-ton operation.

General Description

The milling plant consists of an ore-receiving bin and three steel-constructed buildings housing, respectively, the primary crushing equipment, the secondary crushing equipment, and the ore-storage bins, grinding, flotation, and filtering equipment.

The mill site is the one that was used for a previous copper-concentrating plant, which was operated over a period of years and then dismantled. There remained from this operation some milling equipment, a reservoir for mill-water supply, railroad tracks, tailings-disposal ponds, and a steel-constructed building, all of which were incorporated in the design of the present lead-zinc mill. The location plan of the plant layout is shown in Fig 1.

The flowsheet of the mill is based on the conventional scheme for lead-zinc treatment, however, several innovations were made necessary in order to utilize the existing building. Placing of equipment also was governed by the design of the building; thus making the mill unique in many respects. The following brief description of the mill building as adapted to the flowsheet points out the compactness of the installation and some of its unusual features.

The first floor (Fig 2) houses the grinding equipment, with connecting conveyors for ore feeding, also an air compressor, two vacuum pumps, bins for lead and zinc concentrate storage with individual conveying systems for loading the two kinds of concentrates. Filtering of concentrates is done on the second floor, the concentrates discharging directly into the concentrate-storage bins. On this floor are the lead and zinc filters, two diaphragm pumps, sample-filtering equipment, sample-preparation room and a change room for the mill employees. The flotation machines are on the third floor, at elevations allowing a gravity flow of the lead and zinc concentrates to the thickeners. The annex to the building contains a freight elevator and provides space for all reagent mixing and feeding, also storage of reagents and grinding balls.

All the water used for milling purposes is pumped from the Junction mine and

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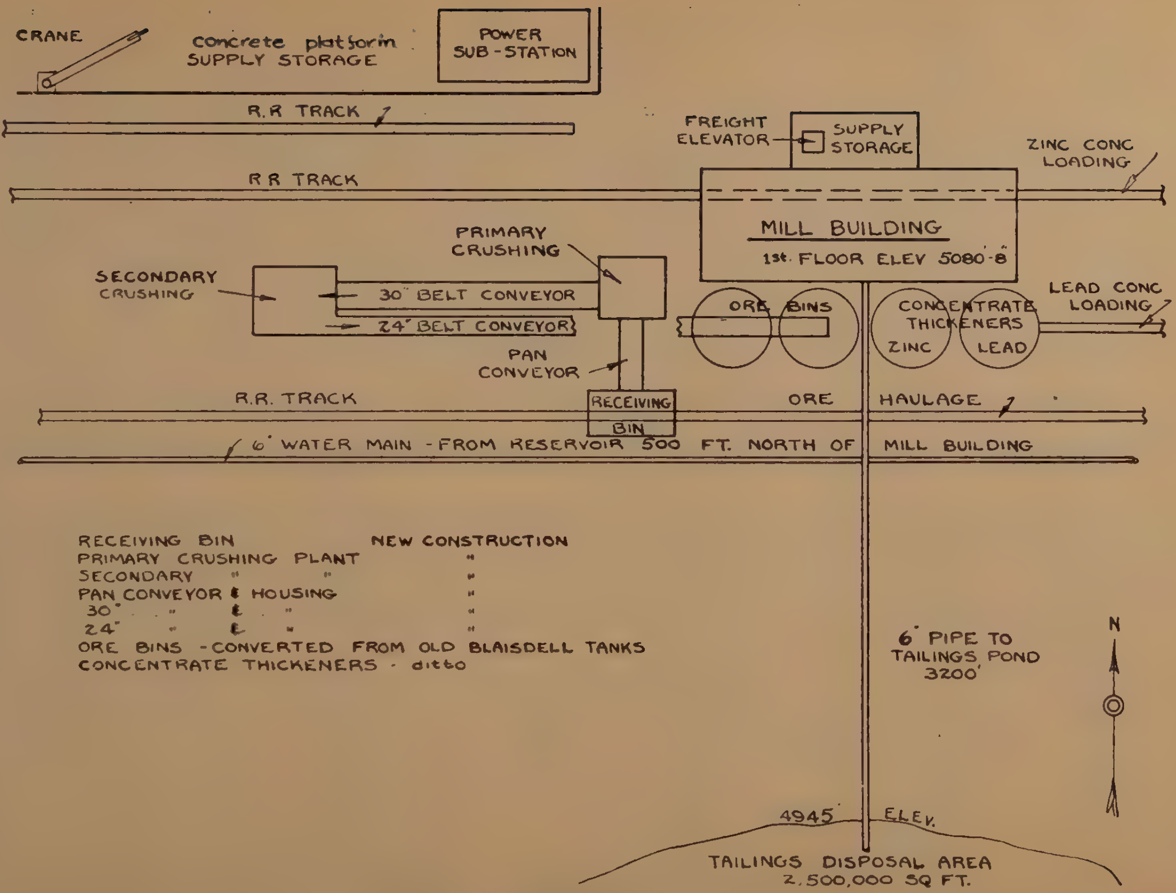

FIG 1—LOCATIONS OF PLANT AND EQUIPMENT.

none of the water used is reclaimed. Power is brought to the mill from the company steam and diesel power plants at 2200 volts and is transformed to 220 and 110 volts for mill use.

ORE TREATED

The lead and zinc generally occur in the ore as galena and sphalerite, while the iron occurs mainly as pyrite. Anglesite and cerussite have been noticed on rare occasions in amounts sufficient to be observed on the pilot tables. The sphalerite varies from a straw-colored variety to a coal-black variety as seen in the ground condition under the microscope, with varying iron content. Of the sphalerite that enters the lead concentrate, the light-colored varieties predominate. The major portion of copper exists as chalcopyrite, with bornite next in frequency of occurrence. Chalcocite is sometimes present in small amounts. Quartz and calcite are the predominating gangue. Ores from various sections of the mines show marked differences in association of minerals and texture. Some ores are so fine grained as to defy liberation of the mineral with any practical grind, and others are fairly coarse grained. In milling, the practice is to grind to approximately 68 pct minus 200-mesh as a compromise to the demands of the various ores. Sphalerite-pyrite locking is generally more severe than that between other minerals. Activation of sphalerite in place by copper salts, tarnished galena surfaces, locking of minerals, and excessive fineness of sphalerite all play a part in reducing the selectivity of the mineral separations.

MILLING PRACTICE

A flowsheet of the mill is shown in Fig 3. It gives a comparatively complete picture of the progress of the ore through the various stages of treatment.

Unloading and Crushing

Ore is received from the mines in gabled-bottom side-dump cars of 50 tons capacity. Loaded cars are weighed on a track scale in the mill yard. The cars are dumped into the receiving bin—a concrete and steel hopper-bottomed bin in an excavation beneath the ore-haulage track. The receiving bin is simply a dumping pocket and was not designed to provide any appreciable ore storage. The ore track accommodates 28 loaded cars above the bin and an equal number of empties below, and serves in lieu of a storage bin for coarse ore.

Ore is transported from the receiving bin by a 34-in. pan conveyor to a steel grizzly having 2-in. spacings. The conveyor is 45 ft between head and tail sprockets and is on an incline of 22° from horizontal. The conveyor speed may be varied between 8 and 14 ft of travel per min. Grizzly oversize is crushed to 2½ in. in an 18 by 36-in. jaw crusher. The crusher operates at 240 rpm and is driven by a 75-hp motor through Texropes. Crushed ore and grizzly undersize are delivered to a 30-in. conveyor belt, which passes under a magnet and discharges onto a 3 by 8-ft vibrating screen, ahead of a 3-ft cone crusher. The cone crusher is driven by a 150-hp motor through Texropes and is set to deliver, with a single pass, a product containing 85 pct minus ½-in. The screen undersize and crusher product are transported by a 24-in. inclined conveyor to an elevated 850-ton ore bin, which originally was one of four old Blaisdell tanks, 29 ft, 6 in. by 11 ft, of reinforced concrete construction. The depth of the tank was increased to 17 ft by adding a ring of 5⁄16-in. plate. A steel-framed housing was erected over the tank and four feeders were installed in the bottom to complete the conversion into an ore bin with a live capacity of 850 tons.

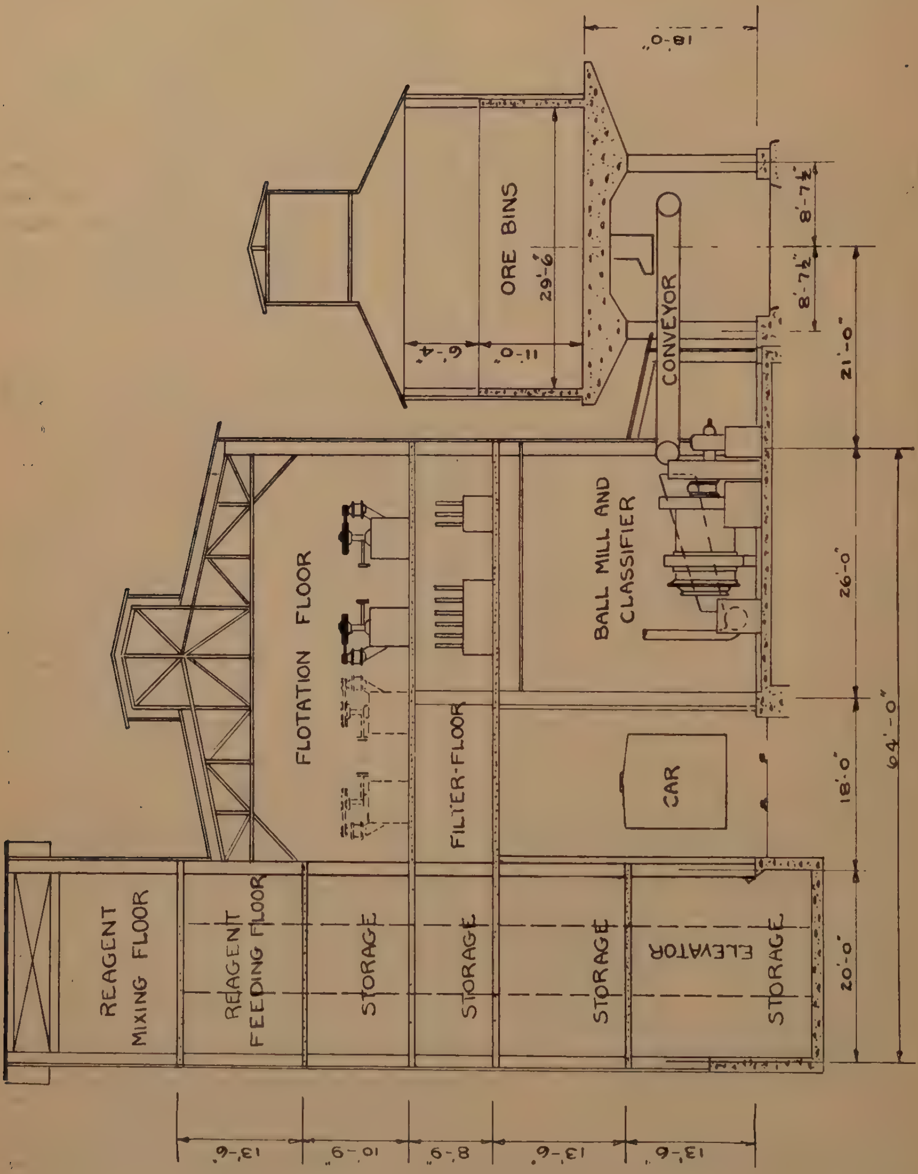

FIG 2—CROSS SECTION OF MILL BUILDING.

Grinding

Ore is drawn from the bin by four conveyor feeders and is delivered to the ball mill by a 20-in. conveyor equipped with a weightometer. The grinding unit consists of a 7 by 7-ft grated-end ball mill operating in closed circuit with an 8 by 23-ft duplex rake classifier. The mill, driven by a 200-hp motor, operates at 23 rpm. Forged steel balls are used, the daily make-up proportioned at 60 pct 2-in. and 40 pct 3-in. balls. The ball load is maintained at a level about 2 in. below the center line of the mill. Ball consumption has averaged 1.23 lb per ton of ore ground. Classifier overflow is maintained between 32 and 35 pct solids with a sand return circulating load around 400 pct. Table 2 shows the grinding product.

Flotation

Classifier overflow is pumped with a 3-in. sand pump to a sampling box ahead of the lead flotation circuit on the third floor of the mill building. The pump operates at 1200 rpm against a head of 45 ft. Fig 3 shows the general arrangement of the flotation flowsheet, in which thirty-two 43 by 43-in. Denver Sub-A cells and one 10 by 10-ft conditioner are employed.

A bank of 12 cells constitutes the lead circuit, the feed entering the No. 4 cell. The circuit consists of a roughing section of six cells, a scavenging section of three cells, concentrate-cleaning and recleaning sections of two and one cells, respectively. Cells No. 4 to No. 9 inclusive produce a rougher concentrate that flows by gravity to the No. 2 cell. Cells No. 2 and No. 3 perform the first cleaning operation and produce a concentrate that is cleaned in the No. 1 cell, where the final concentrate is produced. Cleaner and recleaner tailings return to the rougher section. Concentrate from the scavenger cells No. 10, No. 11 and No. 12 are returned to the rougher section and enter No. 6 cell. The lead tailing from No. 12 cell flows to the conditioner tank, where it is conditioned approximately 13 min and then is split between two parallel banks of 10 cells each, which comprise the zinc flotation circuit.

Feed enters the No. 3 cells of the two zinc banks. Rougher concentrates from cells No. 3 and No. 4 go to No. 1 for cleaning and from No. 5, No. 6, and No. 7 to No. 2 for cleaning. Scavenger concentrate from No. 8, No. 9 and No. 10 returns to the No. 5 cell. The first one or two cells, depending upon the zinc content of the feed, produce finished concentrate, and tailings from these cells join the original feed to the zinc section in the No. 3 cell.

In both circuits all middling products are returned to the different cells by gravity flow in launders. Elevations of the lead banks, conditioner tank and zinc banks of flotation machines permit a gravity flow of pulp through the entire flotation installation and of the final concentrates to the thickeners.

Reagents

A mixture of sodium carbonate, thiocarbanalid and zinc sulphate is added to the ball mill by means of a dry reagent feeder. Additional zinc sulphate, as a 15 pct solution, and sodium cyanide, as a 10 pct solution, are also added to the ball mill. Cresylic acid and sodium ethyl xanthate are added to the classifier overflow as the pulp enters the pump sump. Additions of sodium ethyl xanthate are also made to Nos. 7 and 10 cells in the lead bank. Sodium carbonate is added mainly to produce the desired alkalinity. The pH value of the mill water is 8.2 and lead circuit feed is maintained from 7.6 to 7.8. Thiocarbanalid is used as the collector; a little sodium ethyl xanthate being used intermittently. Thiocarbanalid has proved to be more selective and to pro-

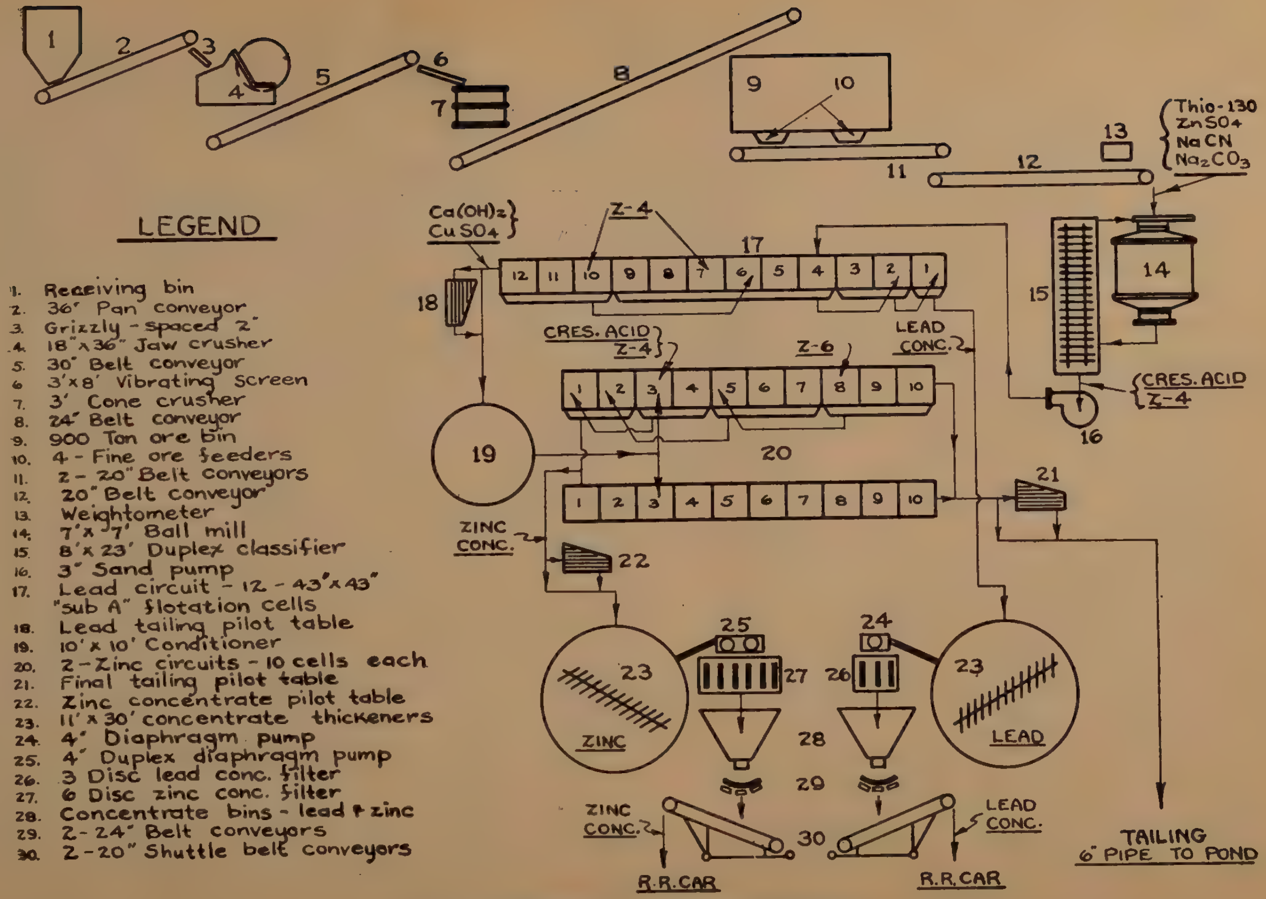
LEGEND
1. Receiving bin
2. 36" Pan conveyor
3. Grizzly - spaced 2"
4. 18"x36" Jaw crusher
5. 30" Belt conveyor
6. 3'x8' Vibrating screen
7. 3' Cone crusher
8. 24" Belt conveyor
9. 900 Ton ore bin
10. 4 - Fine ore feeders
11. 2 - 20" Belt conveyors
12. 20" Belt conveyor
13. Weightometer
14. 7'x 7' Ball mill
15. 8'x 23' Duplex classifier
16. 3" Sand pump
17. Lead circuit - 12 - 43"x 43" "sub A" flotation cells
18. Lead tailing pilot table
19. 10' x 10' Conditioner
20. 2 - Zinc circuits - 10 cells each
21. Final tailing pilot table
22. Zinc concentrate pilot table
23. 11' x 30' concentrate thickeners
24. 4" Diaphragm pump
25. 4" Duplex diaphragm pump
26. 3 Disc lead conc. filter
27. 6 Disc zinc conc. filter
28. Concentrate bins - lead & zinc
29. 2 - 24" Belt conveyors
30. 2 - 20" Shuttle belt conveyors
Thio-130
ZnSO4
NaCN
Na2CO3
Ca(OH)2
CuSO4
Z-4
CRES. ACID
Z-4
Z-6
LEAD CONC.
CRES. ACID
Z-4
ZINC CONC.
ZINC
LEAD
ZINC CONC.
LEAD CONC.
R.R.CAR
R.R.CAR
TAILING
6" PIPE TO POND

FIG 3—FLOWSHEET OF 450-TON LEAD-ZINC MILL.

duce a higher grade of concentrate than xanthate.

Hydrated lime and copper sulphate are added to the lead circuit tailing as it flows to the conditioner. Lime is added by means of a dry feeder but preparations are now underway to change the method to a milk of lime addition. Copper sulphate is added as a saturated solution, overflowing from a lead-lined box. Lime is added in amounts to give a pH of 9.8 to 10.0 in the zinc circuit feed. To accomplish this, 2 to 6 lb is added per ton of original feed, the average being about 3 lb. Cresylic acid and sodium ethyl xanthate are added to the pulp from the conditioner as it enters the No. 3 cell in the zinc circuit. Potassium amyl xanthate is added to the No. 6 cell and at times cresylic acid is added to the No. 6 or No. 7 cell.

TABLE 1—*Metallurgical Data, Nov.* 17, 1945 *to March* 19, 1946

Wet tons of ore milled	52,878.00
Percentage of moisture	1.82
Dry tons of ore milled	51,916.19
Operating days	121
Dry tons milled per operating day	429.06
Dry tons milled per 24 hr of running time	466.45
Kilowatt-hours per ton of ore milled	20.37
Ball consumption per ton of ore milled	1.23
Dry tons lead concentrate produced (smelter receipts)	6,409.029
Dry tons zinc concentrate produced (smelter receipts)	11,614.830
Percentage of moisture in lead filter product	8.0
Percentage of moisture in zinc filter product	9.2
Ratio of concentration lead concentrate	8.1005
Ratio of concentration zinc concentrate	4.4698

PRODUCT ANALYSIS

Product	Ounces per Ton		Per Cent			
	Gold	Silver	Copper	Lead	Zinc	Iron
Heads	0.017	3.42	0.82	6.41	13.60	9.93
Lead concentrate	0.059	15.72	4.49	43.46	9.64	12.99
Zinc concentrate	0.014	3.84	0.79	2.75	50.96	6.28
Tailing	0.010	0.95	0.13	0.66	1.55	10.60

PRODUCT RECOVERY

Lead concentrate	43.62	56.79	67.77	83.70	8.75	16.15
Zinc concentrate	18.73	25.14	21.61	9.60	83.82	14.15
Tailing	37.65	18.07	10.62	6.70	7.43	69.70

Three laboratory concentrating tables are incorporated in the flotation flowsheet; one on lead circuit tailing, one on finished zinc concentrate, and one on flotation final tailing. These pilot tables present at all times a reliable picture of the performance of the different circuits and enable the operator to quickly note changing conditions throughout the plant.

Reagents are mixed on the day shift in quantities adequate for 24 hr of operation. The mixing tanks are piped for air agitation, and are connected to smaller drums on the feeding floor below. Single and double stainless-steel feeders are used for all wet reagent feeding and 18-in. cone-type feeders for the dry reagents. Facilities for close measurement and distribution of the various reagents are very good. Table 4 summarizes information on reagents for operations to date.

TABLE 2—*Typical Screen Analyses of Mill Feed and Grinding Products*
PER CENT OF TOTAL WEIGHT

Screen Size	Mill Feed	Mill Discharge	Classifier Sand	Classifier Overflow
Inch				
0.525	13.75			
Mesh				
4	43.62	12.94		
6	7.46	1.93		
8	4.79	1.40		
10	3.74	1.24		
14	3.76	1.74		
20	2.93	2.19	30.85	
28	2.15	2.98	4.87	
35	1.91	4.98	7.22	
48	1.65	7.80	9.95	
65	1.56	10.97	12.70	4.01
100	1.48	11.95	12.41	6.58
150	1.38	8.47	8.08	10.19
200	1.36	6.47	4.58	12.04
−200	8.46	24.94	9.44	67.18

Concentrate Handling

Finished lead and zinc concentrates flow by gravity to two 29-ft, 6-in. by 11-ft thickeners which were constructed by installing mechanisms and superstructures on two Blaisdell tanks. The underflows from the thickeners, at from 55 to 65 pct solids, are delivered by diaphragm pumps to a three-disk filter for lead and a six-disk filter for zinc concentrate. The lead filter discharges the concentrate directly into a storage bin of 200 tons capacity. The lead concentrate is conveyed from the bin by means of two connecting conveyors, which deliver the material to a car loader when loading box cars, or over the side of gondola cars when that type of car is supplied. The zinc filter discharges concentrate onto an 18-in. conveyor, which delivers the material to a bin for boxcar loading, or to a cross conveyor which discharges directly into a gondola car when that type of car is being loaded. The zinc-concentrate storage bin has a capacity of 400 tons. Vacuum for filtering is normally maintained around 21 in. of mercury by a 16 by 12-in. single-cylinder vacuum pump, converted from an air compressor. The overflows of the concentrate thickeners pass to settling ponds, which in turn produce waste overflows.

TABLE 3—*Screen Assay Analyses, Month Composite*

Screen Size, Mesh	Wt Pct	Assays, Pct			Metal, Pct of Total		
		Copper	Lead	Zinc	Copper	Lead	Zinc
HEADS							
On 65	3.40	0.22	0.74	3.91	1.03	0.40	1.07
100	6.80	0.29	1.37	6.53	2.70	1.47	3.60
150	10.30	0.53	2.05	10.90	7.48	3.34	9.11
200	13.00	0.54	2.68	14.20	9.53	5.50	14.97
Through 200	66.50	0.87	8.50	13.21	79.26	89.29	71.25
	100.00				100.00	100.00	100.00
LEAD CONCENTRATE							
On 65	0.10	15.33	8.85	3.50	0.32	0.03	0.05
100	0.30	15.33	8.85	3.50	0.97	0.05	0.15
150	1.60	10.96	24.00	5.45	3.69	0.72	1.27
200	5.40	7.08	36.29	6.37	8.05	3.66	5.01
Through 200	92.60	4.46	55.27	6.94	86.97	95.55	93.52
	100.00				100.00	100.60	100.00
ZINC CONCENTRATE							
On 65	0.60	1.65	7.88	39.53	1.32	1.70	0.45
100	2.80	0.82	3.35	47.40	3.07	3.38	2.51
150	9.20	0.80	3.45	48.63	9.81	11.44	8.46
200	14.40	0.92	2.96	49.76	17.67	15.36	13.54
Through 200	73.00	0.70	2.59	54.39	68.13	68.12	75.04
	100.00				100.00	100.00	100.00
TAILING							
On 65	4.40	0.15	0.94	2.47	3.93	4.29	8.89
100	8.00	0.19	0.99	1.49	9.05	8.21	9.75
150	11.10	0.24	1.18	1.23	15.83	13.58	11.16
200	12.00	0.19	0.89	1.08	13.57	11.07	10.60
Through 200	64.50	0.15	0.94	1.13	57.62	62.85	59.60
	100.00				100.00	100.00	100.00

Shortly after the start of operations and at the request of the smelters, the method of shipping concentrates was changed from boxcar to gondola shipments. This was the cause of some concern for it was thought that excessive losses in transit would result from the use of open gondola cars, especially with the zinc concentrate, for which the distance to the smelter is 1100 miles.

Gondola shipments have proved entirely satisfactory, however, and now the gondola type of car is actually preferred to the boxcar for concentrate shipping. Of a total of 232 cars of zinc concentrate shipped to date, 191 were gondola cars, and the average loss in transit for all shipments was 189 lb per car.

TABLE 4—*Consumption of Reagents*

Reagent	Lb per Ton Ore Milled	Point of Addition
Sodium cyanide..........	0.67	Ball mill
Zinc sulphate (25.5 pct)...	1.85	Ball mill
Sodium carbonate........	0.36	Ball mill
Thiocarbanalid...........	0.12	Ball mill
Lime (hydrate)..........	3.24	Lead circuit tailing box
Copper sulphate..........	1.78	Lead circuit tailing box
Cresylic acid.............	0.24	Classifier overflow Zinc circuit cells 3 and 7
Sodium ethyl xanthate....	0.12	Classifier overflow Lead circuit cells 7 and 10 Zinc circuit cell 3
Potassium amyl xanthate.	0.05	Zinc circuit cell 6

Tailings Disposal

Approximately two thirds of the total tonnage milled leaves the plant as waste tailing. This tailing pulp flows by gravity through a 6-in. line constructed of transite and wood pipe to an old tailing pond 3200 ft from the mill. The pond has an area of approximately 2,500,000 sq ft and a border about 10 ft high encloses the pond. No further border building will be required for a long time, so the mill tailing is simply spilled into the pond and a clear overflow drawn off through a weir. No tailing water is recovered.

Sampling and Testing

A sample of the crushed ore as it discharges into the storage bin is manually collected at 30-min. intervals. This sample is dried and the moisture content is applied to the mine-shipment scale weights for the dry-ore receipt figure. Weightometer tonnage, as ball-mill daily feed, is adjusted to meet the ore receipts. Shift samples are taken, at 30-min. intervals, of the flotation feed, lead concentrate, zinc concentrate, and mill tailing. The feed and tailing samples are cut by electrically controlled automatic samplers and the concentrate samples are cut by hand. Lead and zinc concentrate shipment samples and car weights supply metal statistics for use in checking mill-production calculations. All sample pulps are prepared for assaying in the sample-preparation room at the mill and the sacked pulps are sent to the mine-assay office for assay.

TABLE 5—*Power Distribution*

Operation	Kw-hr per Ton Ore Milled	Percentage of Total
Unloading and crushing.....	1.42	6.95
Grinding..................	9.79	48.07
Flotation..................	6.60	32.38
Concentrate handling.......	0.95	4.68
Miscellaneous.............	1.61	7.92
Total.....................	20.37	100.00

Dust Control

Every effort is made to provide working conditions that are not injurious to the health of the employees. All chutes junction boxes, and crushing equipment are enclosed and connected to the dust-collecting system. Equipment for the control of dust created in crushing operations consists of a 6500-cfm exhauster and a 7-ft dry collector. Dust settled in the collector is intermittently discharged onto the stream of ore conveyed to the fine-ore bin. In filtering lead concentrate, it developed that a small amount of the material, as fine dust, was carried back into the mill by an uprising current of air through the lead-concentrate storage bin. A hooded enclosure around the filter, with connections to an exhauster, has proved effective in controlling this condition.

In loading concentrates, some dust escapes around the loading area, and makes the matter of clean-up of great importance in minimizing the hazard of contamination by lead dust. The frequent washing of floors is insisted upon; also the wearing of rubberized gloves in performing certain concentrate-loading operations.

Approved respirators are furnished to men engaged in unloading ore, in crushing or in loading concentrate.

Mill Data

In Tables 1 to 5 are shown data covering results obtained from Nov. 17, 1945 to March 19, 1946. Since this four months' period represents the break-in period of a new mill, subsequent operations will no doubt result in additional improvements in metallurgy.

Supplement

Since the preparation of this article, the mill capacity has been doubled and the flowsheet for the zinc circuits changed. Doubling the capacity was effected through duplication of the original installation for fine grinding and flotation, addition of a 75-ft thickener and an 8-ft 6-in. six-disk filter for handling zinc concentrate, and changes in the scheme for handling lead concentrate to provide for the greater tonnage of lead concentrate produced. Operations to date with the increased tonnage indicate a slight improvement in metallurgy. Recoveries of all metals are approximately the same as shown in the preceding data and grade of the zinc concentrate has been raised.

Milling Practices at Concentrator of Morenci Reduction Works

By B. H. Cody,* Member AIME

The Morenci concentrator of the Phelps Dodge Corporation is a part of the project that was completed after five years had been spent in development of the open-pit mine, equipment and process testing in the former No. 6 concentrator, and construction of the reduction works. The reduction works, of which the concentrator is a part, is about a mile and a quarter southeast of Morenci, on the highway that leads up Morenci Canyon.

The concentrator has a nominal capacity for the treatment of 45,000 tons of ore daily. The original plant, which has a rated capacity of 25,000 tons of ore daily and is now in its fifth year of operation, began production on Jan. 30, 1942. The addition, to increase the capacity 80 pct, began operation on Dec. 1, 1943.

The equipment of the original plant has been described in Mining and Metallurgy (May 1942). No changes in equipment were recommended for installation in the concentrator extension and only one major change, forced by a shortage of maintenance labor, has since been made. Metallurgical treatment has been altered in some ways as experiments progressed and as labor and economic conditions dictated.

No complete description of the operation of the concentrator has been published, so this paper will record such operating and equipment changes as have been necessary, the operating methods that have been developed and the metallurgical results that have been obtained. Descriptions are given in the order of the processes with a flowsheet and an equipment list of each process for convenient reference. The questions asked by engineer visitors have guided the detail in which operation and maintenance are described.

During the period Jan. 30, 1942 to Sept. 1, 1946, ore milled has totaled 43,595,253 dry tons, from which 1,643,885 tons of concentrate have been made, containing an indicated production of 783,164,809 lb of copper. The principal results obtained in operation of the concentrator during the years 1942 to 1945, inclusive, and to Sept. 1, 1946 are shown in Table 1. A graphic general flowsheet of the plant is given in Fig 1.

Ore

The Morenci ore that is now being treated is a medium-hard monzonite porphyry in which the principal sulphide minerals are chalcocite and pyrite. A minor amount of covellite is present. Chalcopyrite and bornite have been observed under the microscope.

The chalcocite occurs mainly as a coating on pyrite. Some of the coatings are so thick that the pyrite presents the appearance of being an inclusion in chalcocite. Some are less than one micron thick and are barely visible at 500 diameters magnification of a polished cross section. Some chalcocite is present as a dissemination in the gangue.

Various oxidized copper minerals are present as a result of oxidation in place but they represent a minor proportion of the copper minerals of the ore body. Oxidation

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of some of the copper in ore near old stoping areas is relatively high but is not large in quantity. The bulk of the oxidized copper is formed subsequent to the breaking of the ore, particularly during the simultaneous occurrence of heat and humidity. Unaltered ore contains about 5 pct of the copper in an oxidized condition, mainly as imperceptible coatings on the chalcocite. Additional oxidation by weathering begins when the ore is broken and continues until the ore is delivered to the crushing plants. Little additional oxidation occurs in crushing because of the short time of residence. The higher temperatures produced in the wet-grinding process then cause a further increase in oxidation, until the proportion is nearly double the original oxidized copper content.

TABLE 1—*Summary of Operating Results*

Data	1942	1943	1944	1945	1946 to Sept. 1
Dry tons ore treated	7,592,246	9,577,595	11,265,526	10,390,862	4,769,024
Percentage of moisture	2.91	2.74	3.19	3.46	3.18
Dry tons per operating day	22,663	27,132	35,992	33,957	39,413
Dry tons per mill per 24 hr	1,719	1,700	1,639	1,617	1,792
Assays:					
Feed, pct:					
Total copper	0.84	0.97	1.11	1.15	1.18
Oxide copper	0.07	0.08	0.09	0.11	0.11
Iron	3.5	3.5	3.5	3.5	3.3
Sulphur	3.5	3.2	3.1	3.07	3.00
Pyrite	6.19	5.57	5.32	5.26	5.11
Concentrate, pct:					
Copper	20.53	21.38	23.79	26.81	27.70
Insoluble	5.4	5.4	6.4	5.5	6.4
Iron	31.3	30.1	28.8	27.9	26.4
Moisture	9.30	9.15	9.43	9.52	9.61
Tailing, pct:					
Total copper	0.12	0.14	0.16	0.18	0.19
Sulphide copper	0.05	0.06	0.07	0.07	0.08
Ratio of concentration	28.59	25.53	24.95	27.34	27.80
Extractions, pct:					
Total copper	85.79	86.37	86.86	85.17	84.68
Sulphide copper	93.30	93.90	93.72	93.54	93.04
Iron	31.1	33.3	32.6	28.8	28.9
Pyrite	38.1	46.6	46.8	41.5	41.3
Labor:					
Man-shifts per day	242.1	267.9	390.5	404.2	459.0
Dry tons treated per man-shift	93.6	101.3	92.2	84.0	85.9
Fresh water:					
Gallons per ton ore	219	201	213	218	179
Power, kw-hr:					
Crushing	0.92	0.87	0.97	1.04	1.05
Grinding	8.40	8.61	8.76	8.81	8.06
Flotation	2.03	2.11	2.22	2.37	1.99
Miscellaneous	2.21	2.25	2.59	2.59	2.48
Total	13.56	13.84	15.54	14.81	13.58
Reagents, lb per ton ore:					
Chemicals	0.033	0.037	0.033	0.041	0.037
Oils	0.079	0.082	0.115	0.114	0.098
Lime	3.327	3.359	3.500	3.624	3.896
Grinding mill steel, lb per ton ore:					
Forged balls	1.37	1.41	1.49	1.54	1.36
Cast iron balls	0.18	0.13	0.15	0.15	0.25
Liners	0.17	0.17	0.12	0.17	0.13

The gangue varies from a soft type in which the feldspars are highly altered to hard ore in which no alteration has occurred. As delivered at the concentrator the ore varies in iron content from 2.5 to 5 pct and the copper content from 0.85 to 1.55 pct.

The average total copper content of the ore milled during the period Jan. 30, 1942 to Sept. 1, 1946 has been 1.048 pct, of which 0.092 pct copper was present as oxidized minerals. The analysis of the ore milled during 1945 is as follows:

	Per Cent
Total copper	1.15
Oxide copper	0.11
Sulphide copper	1.04
Silica	65.2
Alumina	17.9
Iron	3.5
Lime	0.2
Sulphur	3.07
	Oz per Ton
Gold	0.0013
Silver	0.051

The calculated sulphide-mineral contents of the ore are 1.30 pct chalcocite and 5.26 pct pyrite. The weight of pyrite is more than four times the amount of chalcocite. The proportion of oxidized copper is 9.57 pct of the total copper.

Primary Crushing

The flowsheet of the primary crushing process is shown in Fig 2.

Mine cars contain an average of 85 dry tons of ore. Dumping is against an ore pocket above the grizzly. It is specified to be at the rate of one car every 1½ min. A protected concrete wall supports the railroad bridge and divides the ore flow to both sections of the grizzly and to the two sides of the crusher.

No delays of consequence have occurred as a result of coarse rock lodging above the grizzly. The ore contains approximately 3.5 pct moisture and, except when the moisture content is above 5 or 6 pct, the angle of repose in the dumping pocket lies between 40° and 50°. The angle of repose may become as high as 75° when wet ore is received. Alternate receipt of coarse and fine rock then assists in clearing the pocket. Provision was made in the design for the installation of chain feeders above the grizzly, with the object of regulating the rate of feed to the gyratory crusher and increasing the efficiency of the grizzly, but the necessity for them has not arisen. Choking of the gyratory crusher by large rock occurs but jams are easily broken by the use of a collapsible hook attached to a 10-ton traveling crane immediately above the crusher, beneath the top-floor hatch covers. The ore travels across the grizzly rapidly. About 25 pct of the material passes through the openings and this constitutes a major portion of the undersize in the run-of-mine ore. The crusher has ample capacity, so high efficiency of the grizzly is not necessary.

The 60-in. gyratory crusher is driven by a 500-hp wound-rotor induction motor. The motor has a rated speed of 265 rpm. For the amount of ore that has been crushed, liner replacements in the crusher have been few. After 24 million tons of ore had been crushed, the lower tier of the lower concave liners was replaced; after 42 million tons of ore, 11 liners of the upper tier of the lower concaves and 9 liners of the lower tier of the upper concaves were replaced in the sides of the crusher bowl, where principal impact of the ore occurs. The concave liners under the crusher spider do not receive heavy wear because of the diversion of the ore by the dumping-pocket dividing wall and the spider. Replacement of the mantle liners is scheduled for the near future.

The repair facilities provided have aided the efficiency of the work. The eccentric can be lowered for inspection and replaced in 8 to 10 hr by means of the hydraulic jack. The hook of the 100-ton crane can be lowered through hatchways to all floors of the crusher building to handle heavy parts. A freight elevator of 5 tons capacity is useful for the transportation of small supplies and workmen.

The gravity-feed lubrication system provided for the gyratory crusher has performed satisfactorily. The amount of oil in use in the system is about 3500 gal. The lubricant is a straight mineral oil of 30 SAE made from napthenic stock. It has the following properties: flash point, 360°F; fire point, 410°F; Saybolt viscosity, 515 sec. at 100°F. All cleaning of the oil is done about three times each month by

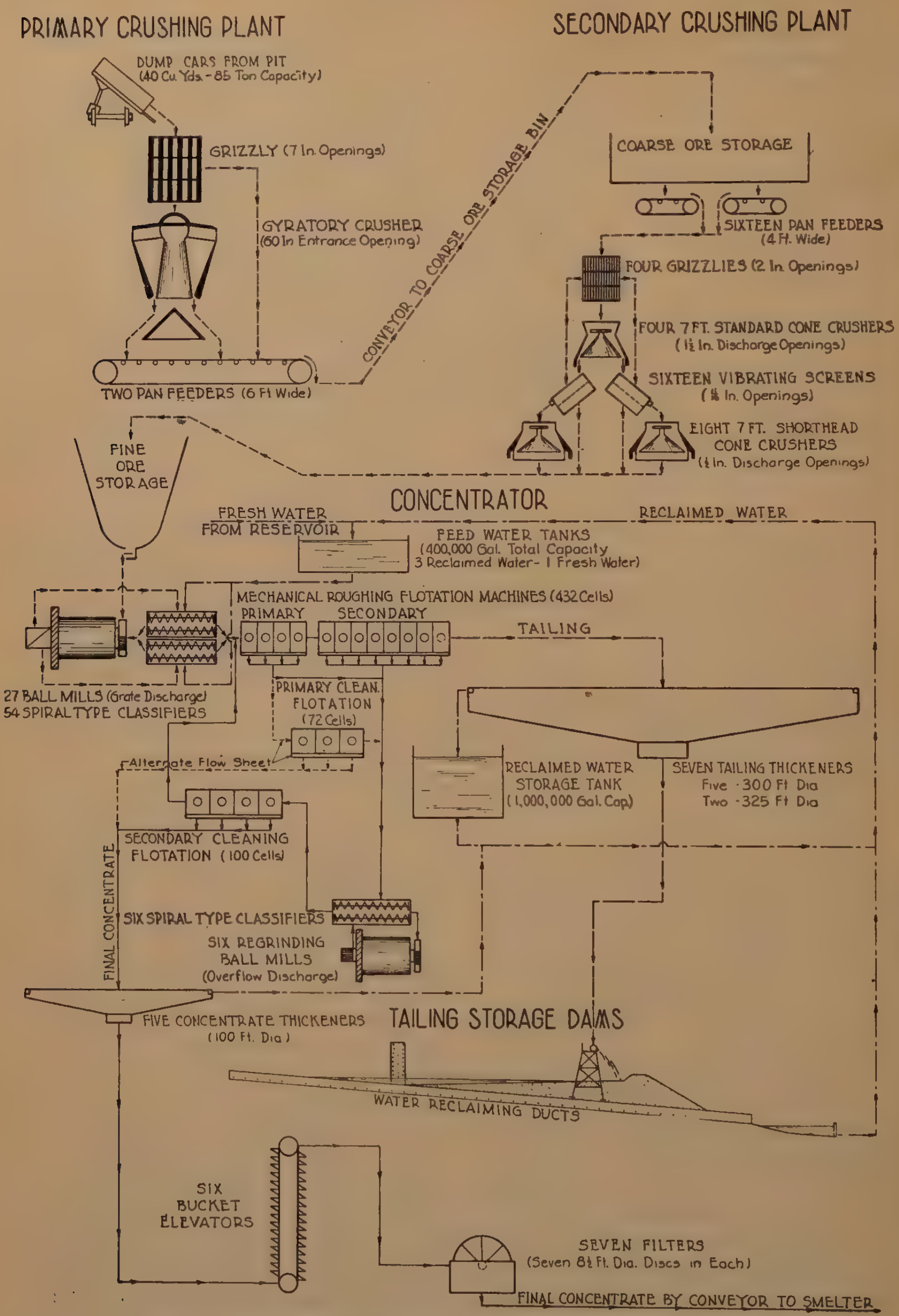
PRIMARY CRUSHING PLANT
SECONDARY CRUSHING PLANT
DUMP CARS FROM PIT
(40 Cu. Yds. - 85 Ton Capacity)
GRIZZLY (7 In. Openings)
GYRATORY CRUSHER
(60 In. Entrance Opening)
TWO PAN FEEDERS (6 Ft Wide)
CONVEYOR TO COARSE ORE STORAGE BIN
COARSE ORE STORAGE
SIXTEEN PAN FEEDERS
(4 Ft. Wide)
FOUR GRIZZLIES (2 In. Openings)
FOUR 7 FT. STANDARD CONE CRUSHERS
(1½ In. Discharge Openings)
SIXTEEN VIBRATING SCREENS
(⅝ In. Openings)
EIGHT 7 FT. SHORTHEAD CONE CRUSHERS
(½ In. Discharge Openings)
FINE ORE STORAGE
CONCENTRATOR
FRESH WATER FROM RESERVOIR
RECLAIMED WATER
FEED WATER TANKS
(400,000 Gal. Total Capacity
3 Reclaimed Water - 1 Fresh Water)
MECHANICAL ROUGHING FLOTATION MACHINES (432 Cells)
PRIMARY
SECONDARY
TAILING
27 BALL MILLS (Grate Discharge)
54 SPIRAL TYPE CLASSIFIERS
PRIMARY CLEAN. FLOTATION
(72 Cells)
Alternate Flow Sheet
RECLAIMED WATER STORAGE TANK
(1,000,000 Gal. Cap.)
SEVEN TAILING THICKENERS
Five - 300 Ft Dia
Two - 325 Ft Dia
SECONDARY CLEANING FLOTATION (100 Cells)
FINAL CONCENTRATE
SIX SPIRAL TYPE CLASSIFIERS
SIX REGRINDING BALL MILLS
(Overflow Discharge)
FIVE CONCENTRATE THICKENERS
(100 Ft. Dia.)
TAILING STORAGE DAMS
WATER RECLAIMING DUCTS
SIX BUCKET ELEVATORS
SEVEN FILTERS
(Seven 8½ Ft. Dia. Discs in Each)
FINAL CONCENTRATE BY CONVEYOR TO SMELTER

Fig 1—Concentrator flowsheet, Morenci Reduction Works.

centrifuging. The oil is heated to temperatures between 90° and 100°F for feeding to the crusher at flows varying from 25 to 35 gpm. Separate lubrication is provided for the crusher spider.

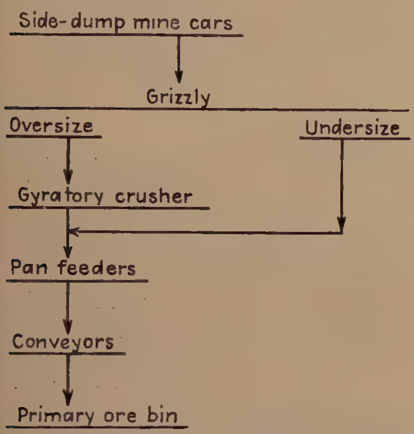

FIG 2—FLOWSHEET OF PRIMARY CRUSHING.

Side-dump cars have capacities of 40 and 43 cu yd.

Grizzly is manganese dressed, 40° inclination, 40 by 16 ft area, 7-in. openings, 35 pct open area.

There is one 60-in. gyratory crusher; two pan feeders, 72 in. wide, 38 ft between shafts; two 54-in. 10-ply-belting conveyors running at 475 ft per min., 15° inclination.

The ore bin holds 23,000 tons (nominal).

Power consumption by the gyratory crusher and the lubrication system has been low (Table 2). The average amount of ore crushed by the primary crushing plant per hour of operation is shown also in Table 2.

TABLE 2—*Power Consumption*

Year	Kw-hr per Ton Ore Milled	Percentage of Total Concentrator Consumption	Tons per Hour Crushed
1942	0.06	0.47	3,773
1943	0.05	0.38	3,375
1944	0.05	0.36	3,421
1945	0.04	0.28	3,230

The lower average for the year 1945 was caused by time for repairing the conveyor belt, when the crushing rate was limited to the capacity of one conveyor.

Feeding of the conveyors is regulated by the speed of the pan feeders. These feeders are always shut down with an ore load upon them, in order to seal the dust-control system for proper operation. Though the pan feeders are equipped with a variable-speed drive up to 39 ft per minute, they are set permanently to operate at their lowest speed of 30 fpm.

The manganese-steel chain links and rollers originally provided on the pan feeders were too light, and breakage caused considerable interruption of production. The links were redesigned by the manufacturer to apply additional metal where possible without alteration of the design of the entire feeder. The two-piece rollers were replaced by rollers built in one piece. The strengthening of the links and the new design of the rollers apparently have been successful in eliminating interruptions of operation due to chain breakage.

The spill conveyors with which the pan feeders are equipped have been worth the installation cost by saving labor necessary for clean-up work. Auxiliary conveyors to deliver spillage to the main conveyor farthest from the pan feeders have been installed since operation of the plant began.

The pan feeders discharge into ore pockets from which the ore flows across finger grizzlies and drops about 18 in. to the conveyor belts. Impact idlers are located under the loading points. The loading arrangement has protected the belts from cutting at the loading point but there has been some damage to the belts caused by ore jams, which form in the discharge chutes of the tripper cars when wet ore is received. The recent installation of iron detectors on these two conveyors decreases the damage due to tramp iron without causing many interruptions of crushing-plant operation. Though the belts have vulcanized splices, the inferior quality of the materials available because of the shortage of natural rubber products for splicing has caused an abnormally large number of splice failures. During the repair of a conveyor belt the production rate of the primary crushing plant is limited to the capacity of one conveyor. The conveyors are equipped with roller-type holdbacks, which have been reliable and have required little maintenance.

The tripper cars have an unusually long wheelbase (25 ft) to minimize conveyor-belt misalignment caused by weaving of the tripper cars upon the rails. They are equipped with solenoid-operated motor brakes.

The primary crushing plant operates on a 24-hr schedule except when a day shift that is convenient to both pit and concentrator operations is utilized for minor repair work. Major repairs are made if possible when the plants are not scheduled to operate. Operation of the primary crushing plant is done by four men on each shift, exclusive of the dumping of the ore by the train crew. One man is stationed at the control board on the top, or dumping, floor. He can start or stop the gyratory crusher and can stop any major equipment used in the primary crushing process. An electrical horn-signal system connects principal locations, through which operators can be informed by the dumping-floor operator of the arrival or departure of trains. All equipment except that for lubrication or dust control is shut down when no ore train is waiting to enter the crushing plant. This causes some wear of switch gear but saving in power and decreased wear of other equipment more than compensates for the extra maintenance of control equipment.

A second operator is stationed on the fourth, or crusher drive, floor, where he observes the crusher discharge and the amount of ore in the surge pockets above the pan feeders. When the load in the pockets is at the correct stage he signals the operator on the dumping floor that a carload can be dumped. If delays in dumping occur he can start and stop operation of the pan feeders on the second floor. He notes the flow of lubricating oil and its temperature and can observe the crusher power requirement, which is shown by ammeters in the electrical control room on that floor.

A crusher operator-helper is stationed on the second, or pan-feeder, floor. His duties are the observation of the operation of the pan feeders and of the feed end of the conveyor belts.

A tripper-car operator works at the discharges of the conveyors in the superstructure of the primary ore bin on distribution of the ore in the bin. It has not been possible to make uniform distribution of the ore by continuous travel of the tripper car. Repair delays in the secondary crushing plant cause irregular withdrawal of ore from the bins, thus leaving high and low spots in the ore bed. Starting and stopping of the conveyors and moving of the tripper cars is controlled by this operator.

All operating floors within the primary crusher building are kept under low pressure by a washed-air ventilating system and the crushing spaces are under low suction from the dust-control system. As a consequence, the operators' stations are relatively free of dust, and atmospheric conditions in the building, particularly in warm weather, are excellent. A dust-sealing curtain planned for the dumping pocket was not installed because material was unavailable. It is not now necessary because sufficient flow of air into the dumping pocket is produced by the dust-control system to prevent much dust from escaping.

Secondary Crushing

The secondary crushing plant is unusual in the very compact arrangement of the equipment (Fig 3). It is divided into two principal sections, each including two complete crushing units. One section of two units prepares and delivers ore to the original part of the concentrator. The other section delivers crushed ore to the concentrator extension. One of the two units of this section can deliver ore to either part of the concentrator. Each secondary crushing unit includes the following major items of equipment: pan feeders, 4; feeder conveyor, 60-in. belt, 1; grizzly, 1; standard cone crusher, 1; vibrating screens, 4; shorthead cone crushers, 2.

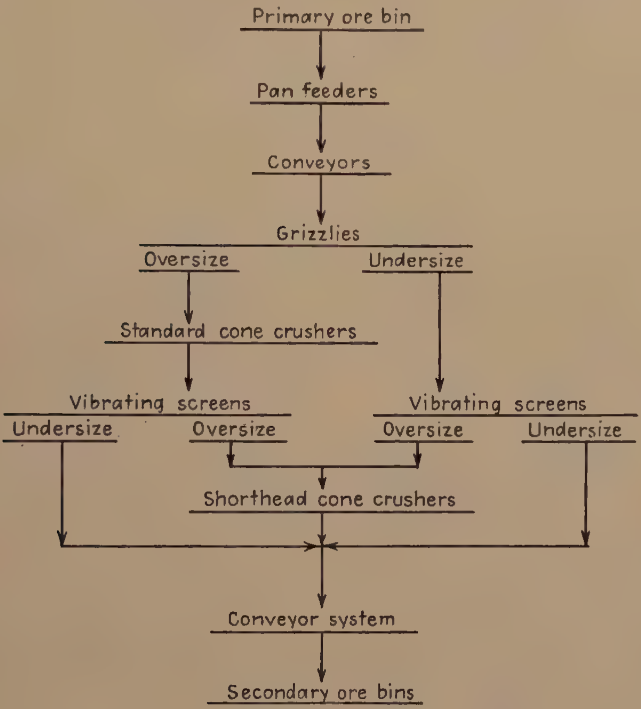

FIG 3—FLOWSHEET OF SECONDARY CRUSHING.

The ore bin holds 23,000 tons (nominal).

There are 16 pan feeders, 48 in. wide, 13 ft between shafts; four 60-in. belt conveyors; four grizzlies, each 4½ by 11 ft, made of manganese-steel bars with 2-in. openings, 40° inclination and four standard 7-ft cone crushers.

There are 16 deck vibrating screens, 5 by 10 ft; 1¹⁄₁₆-in. openings, 25° inclination, and eight 7-ft shorthead cone crushers.

Two series of 54-inch belts.

Suspended type 29,000 tons (nominal).

The discharges of the pan feeders pass over finger grizzlies before dropping on the conveyor belts, which are supported at the loading point by impact idlers. The 10-in. size of ore is discharged by the conveyors into ore pockets preceding the grizzlies and the grizzly oversize product passes directly into the standard cone crushers. Two worn manganese-steel ball-mill shell liners are swung over each standard cone crusher to divert the grizzly product vertically downward to the crusher feed plate.

The grizzlies were installed to replace four 6 by 14-ft rod, deck vibrating screens, which were equipped with 1½-in. diameter rods having 2-in. spaces between. Though the screens were more efficient than the grizzlies and fewer rock jams occurred upon them, the amount of maintenance labor required to keep them in operation was more than was available in the period of labor shortage during World War II. The spacing between the manganese-steel bars of the grizzlies tapers from 2½ in. at the bottom to 2 in. at the top and the deck has an inclination of 40°. The continuity of operation and ease of maintenance of the grizzlies has been satisfactory and the standard cone crushers have been able to handle the oversize product, so it is doubtful whether the screens will be returned to service. This has been the only major change of equipment of the Morenci concentrator.

The 7-ft standard cone crushers are set to an opening of 1½ in. They crush the 10-in. feed to a nominal 2-in. product size, which is discharged upon two 5 by 10-ft vibrating screens set at an inclination of 25°. The grizzly undersize discharges onto two similar screens. The screen decks are equipped with rods ½ in. in diameter by 29 in. long, set across the screen in two panels with 11/16-in. openings between the rods. Changing of rods is not difficult. They are mounted in rubber spacers to make assemblies of 10 rods each. The assemblies are held under a central member of the screen bed and clamped at the outside edge of the bed. A support along the center line of each half of the screen causes the bed to be slightly arched. Screening action is efficient on ore containing less than 6 pct moisture, and the vibratory action is strong enough to move wet ore over the screen bed.

The undersize of the screens is divided to feed two 7-ft shorthead cone crushers, which are set at openings of ⅜ to ½ in. As the hardness of the ore varies considerably, the frequency of adjustment varies from one to five 24-hr periods. All cone crushers are driven at 446 rpm by 300-hp, 900 rpm synchronous motors through 18 E-section V-belts.

Two systems of 54-in. conveyors deliver the ore to the two secondary ore bins of the concentrator. One conveyor of each series is equipped with a stationary magnet, which collects tramp iron that escapes from the secondary crushing plant. These magnets have been useful in decreasing damage to the feeder belts of the grinding section. One of each series of conveyors is also equipped with a conveyor scale having a remote indicator at the operator's stand of the shorthead cone crusher.

Inclined portions of all conveyors of the secondary crushing plant are built on a slope of 20°. The relatively steep slope causes some spillage at the feeding points until the load has increased to normal but when uniform operating conditions have been established the amount of spill is not great. All inclined conveyors are equipped with roller-type holdbacks.

There are two gravity-feed lubrication systems, each providing lubricant for the crushers of two secondary crushing units. A supply of about 10,000 gal of lubricant is maintained in each system. The flow to each cone crusher varies from 40 to 50 gpm at temperatures of 90° to 100°F. The same type of oil is used to lubricate the cone crushers that is used for the gyratory crusher. Centrifuge cleaning of the oil is done on alternate days for each system. One man is assigned continuously to that duty.

The secondary crushing plant has no definite schedule of operation. Operating times of the various units are governed by repairs that are necessary and ore supply or capacity available in the primary and secondary ore bins. Operation of one or more units is in progress most of 24 hr daily.

Seven men on each shift are required for operation of the secondary crushing plant. Two men observe the operation of the feeder conveyor belts, the grizzlies and the standard cone crushers. They can stop and start the conveyor belts when grizzly or crusher chokeups occur. Two men operate the shorthead crushers and screens. Control panels for each of the four secondary crushing units are on the shorthead crusher floor and by these the operator of the shorthead cone crusher can vary the speed of the pan feeders and the stopping or starting of all equipment of two crushing units. Ammeters showing the power requirements of all the cone crushers of two units are visible from the operator's station. He also observes the loading of the discharge conveyors as shown by the remote indicator connected with the conveyor scales. Two men operate the tripper cars to distribute the crushed ore in the secondary ore bins. One man attends to the gravity-fed lubrication system and acts as an extra operator. A shift boss, who is responsible to the concentrator shift foreman only for ore supply, supervises the operation of both primary and secondary crushing plants.

The unit power consumed by the secondary crushing process is composed of the items shown in Table 3.

TABLE 3—*Power Consumption, Secondary Crushing*

Year	Kw-hr per Ton Ore Milled			Percentage of Total Concentrator Consumption		
	Crushers	Conveyors	Miscellaneous	Crushers	Conveyors	Miscellaneous
1942	0.37	0.39	0.10	2.68	2.88	0.75
1943	0.33	0.36	0.13	2.42	2.61	0.90
1944	0.39	0.39	0.14	2.71	2.69	0.90
1945	0.43	0.42	0.15	2.88	2.85	1.02

Crusher power includes only the amounts consumed by the cone crushers and their lubrication systems. Conveying power is that consumed by all conveyors and pan feeders of both the primary and secondary crushing plants. Power consumption of screening, dust control, shop and lighting of the crushing plants is included under Miscellaneous.

A repair area is provided at one end of the secondary crushing plant. Two 30-ton cranes serve the shop and crushing plant.

The screen analysis of the product of the secondary crushing plant is shown in Table 4 together with the average crushing rate of each of the four units of the plant.

TABLE 4—*Screen Analysis and Crushing Rate*

Screen Analysis			Year	Crushing Rate, Tons per Hour
Screen Size	Percentage of Each Size			
	Individual	Cumulative		
Inch:			1942	869
On 1.050	14.5	14.5	1943	925
0.525	43.2	57.7	1944	828
Mesh:			1945	726
On 4	27.2	84.9		
65	10.2	95.1		
200	2.1	97.2		
Through 200	2.8	100.0		

The average for the four years is 826 tons per unit hour. The decrease in the year 1945 was caused by operation of incomplete units because of the shortage of maintenance labor.

The four principal dust-control systems of the secondary crushing plant and the single unit of the primary crushing plant are of the same type. Following the hooding at the dust-disseminating locations, and connected with them by the air ducts, is a cyclonic dry-dust collector under suction from a fan that discharges into a cylindrical spray chamber in which cyclonic action causes the impingement of the finer dust particles upon a water-washed interior labyrinth of walls. Dry dust is returned through trap valves to a conveyor. The wet sludge from almost all the systems flows to a pumping unit, by which it is forced through piping to several classifiers of the grinding section.

The collecting hoods are designed to spread the air intake over a large area by the use of conical or pyramidal tops. The entrance of the air into the ducts is at a sufficient distance from the point of greatest turbulence so that only the smaller dust particles enter the ducts. Enough air is drawn into the hoods to prevent exit of coarse or fine dust particles.

By limiting the size of the dust particles that enter the ducts, the wear resulting from transportation of the dust is kept at the minimum and permits the use of low air velocities in the ducts. The low velocity assists in decreasing wear and keeps the static resistance of the duct systems at the minimum. Notwithstanding these precautions, wear occurs at bends or at points of turbulence and resistance to wear at those places is increased by the application of a plastic rubber compound, which has been successfully used also in protecting fan-impeller blades.

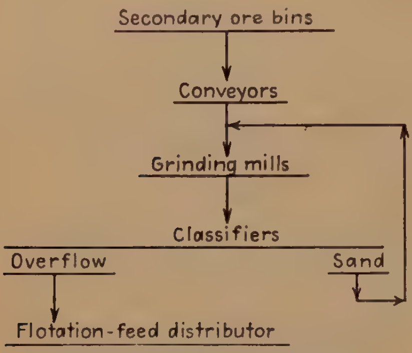

Fig 4—Grinding flowsheet.

The ore bins are suspended type, 29,000 tons (nominal).

There are 27 feeder conveyors, 36 in. wide, and 27 weighing conveyors, 24 in. wide.

The 27 ball mills are grate type 10 by 10½ ft., and the 54 duplex 54-in. classifiers are submerged-spiral type.

Tanks are rectangular, 368 and 510 cu ft, respectively.

Smaller dust-control units are placed at isolated dust-dissemination points. They may include both dry and wet collectors or wet collectors only. The total volumetric capacity of all the dust control systems of the concentrator is 143,550 cu ft per minute.

Clean-up of ore spills by shoveling unavoidably results in the production of dust in the air, but sweeping is avoided where possible by the use of a semipermanent vacuum-cleaning system. A large portable vacuum cleaner was provided for the crushing plants but transportation of the unit required so much manipulation that time available for cleaning was short. By installing the vacuum cleaner in a permanent location, where it was connected to a vacuum piping system equipped with self-closing inlet valves, the use of the system required only the transportation of the vacuum-cleaning nozzles and hoses. The time available for cleaning was increased and a steadier vacuum was secured. Installation of a stationary large-capacity unit to replace the portable vacuum cleaner is planned.

Maintenance, repair, testing and improvement of the dust-control units are done under the direction of a dust-control engineer, an assistant, a repair mechanic and a helper. The mechanical crew is assigned to continuous duty on maintenance, repair and minor alterations of all the dust-control systems. Dust counts are made periodically and systematic efficiency tests record the performance of the units. Regular inspections are made weekly by the dust-control engineer in company with the repair mechanic.

Grinding

Except for the return of reclaimed water, the concentration processes beginning with grinding (Fig 4) are divided into two separate plants having equipment in proportion to the 16 grinding mills of the original plant and the 11 mills of the extension. Each of the 27 grinding units consists of one grinding mill and two classifiers as principal equipment. Ore for each unit is drawn from the bin through one or more of four horizontal feeder gates and is discharged upon a feeder conveyor driven by a variable-speed direct-current motor. From the feeder conveyor it is transferred to a weighing conveyor, which delivers the ore to the grinding-mill scoop housing. All new feed to the grinding units and the circulating sand load from the classifiers enters the grinding mill through the double-entrance scoop. The central drum-feeder portion of the scoop is used only for the addition of grinding balls.

Conveyor scales weigh the feed to the grinding mills and when set to a desired rate of feed regulate the speed of the feeder conveyors through a series of relays. The scale is electrical in principle and is dependent upon frictional contact at only one point, where an idler resting upon the return portion of the conveyor belt transmits the belt speed effect to the scale head through synchronously acting motors. The scales have met all requirements of accuracy and can be easily tested and repaired.

In addition to a conveyor scale for each grinding mill, watt meters are available for most mills. Meter sockets have been provided so that meters can be plugged in to determine the power consumption. Accurate capacity and power-consumption tests can therefore be made on any mill. The ball-mill shells are 10 ft in diameter by 10½ ft long inside the liners and are driven at 18 rpm by 800-hp, 257-rpm synchronous motors.

The grinding-mill discharge flows from the trunnion level through a launder built on a slope of ¾ in. per foot to the classifier feed entrance. Enough water is added at the mill discharge to dilute it to not less than 50 pct solids. Classification of sand and gravel occurs in the launder when the solid content is less than that figure. In order to repulp the feed for classification, the grinding-mill discharge is fed to the classifier upon a stream of water added from a jacket in the corner of the turn of the launder. Turning vanes in the launder at the classifier feed entrance introduce the feed at 90° to the pool. Hydraulic classification traps in the launders near the mill-discharge receiving box remove scrap balls from the grinding circuit.

The operating characteristics of the submerged-spiral type of classifier have been satisfactory. The 54-in. diameter duplex spirals operate at 5 rpm and little agitation is evident in the pool at that speed. The area of the pool to the point where the sand emerges is 192 sq ft and the maximum vertical depth is 6 ft 10 in. The classifiers are set on a slope of 3⅜ in. per foot. The sand load of the spirals is 400 to 500 pct of the weight of the new feed.

The grinding-mill feed is nominally ¾ in. in size but varies from ½ to 1 in. in maximum size of ore particle. The product of the grinding units is kept at about 4 pct on 65-mesh or 16 pct on 100-mesh, which yields about 62 pct of sizes smaller than 200-mesh. Periodic screen tests and determination of the solid content of the classifier overflows by the operators keep the product size close to the required figure. Flow of lime hydrate to each grinding mill is controlled by the operator, to keep the grinding circuit alkaline and the alkalinity at the correct point for flotation. The grinding bosses regulate the addition of collector reagent to each grinding mill to a specific figure.

One man operates four grinding units under the direction of an assistant shift boss, who is in charge of either 11 or 16 grinding units, according to the section of the concentrator in which he works. Two employees attend and regulate the two lubrication systems of the grinding mills. This permits the operators to give complete attention to the problems of their work. Nine workmen and two bosses are therefore required on each shift for capacity operation.

The production of the grinding units has varied considerably. In addition to changes of the ore in regard to grindability or response to classification, labor efficiency, feed screen analysis, mill ball loads and interruptions of operation for repair work, have all had either adverse or beneficial effects.

Average grinding data are shown in Table 5. Power figures include amounts required by feeder conveyors, grinding and classification.

TABLE 5—*Average Grinding Data*

Year	Feed, Pct		Product, Pct		Tons per Mill per Day	Kw-hr per Ton	
	On 0.575 In.	−200 Mesh	On 65 Mesh	−200 Mesh		Feed	−200 M Prod.
1942	34.3	6.9	4.0	63.0	1,719	8.40	14.97
1943	46.8	4.6	4.0	62.5	1,700	8.60	14.87
1944	45.6	4.0	4.0	62.3	1,639	8.76	15.03
1945	57.6	2.8	3.4	64.1	1,617	8.81	14.37
1946 to Aug.	46.2	4.1	4.1	62.1	1,772	8.13	14.01

During July 1946 an average unit production of 1846 tons per 24 hr was obtained from each mill from feed containing 32.8 pct of plus ½-in. sizes with a power consumption of 7.77 kw-hr per ton of ore milled.

During the war years it was necessary to employ women to operate the grinding sections. The work was heavier than their physical capabilities would enable them to perform and the principles of the grinding operation were beyond the understanding of most of them. Any efforts made to maintain the production per unit at the maximum possible usually resulted in overloading and spillage. The unit production rate has shown progressive increase in 1946 as more male labor has become available for operation and maintenance of crushing and grinding equipment.

Forged heat-treated alloy-steel balls of 3-in. diameter are used in the grinding mills. The use of 2½-in. balls is being tested in four mills. Results to Aug. 1, 1946 are compared in Table 6 with those from two mills to which 3-in. balls are being added. The test had been in progress for more than four months when these data were assembled. For the same tonnage ground per day to the usual fineness the 2½-in. balls produced finer grinding. An extensive investigation of ball rationing is contemplated. Consumption of alloy-steel balls has averaged 1.5 lb per ton of ore ground during the years 1942 to 1945, inclusive.

TABLE 6—*Efficiency of Balls*

Operation	2½-inch Balls	3-inch Balls
Dry tons milled per 24 hr.......	1,760	1,769
Kw-hr per ton.................	7.84	7.88
Tons ground through 65-mesh per 24 hr..........................	1,575	1,564
Tons ground through 200-mesh per 24 hr.....................	1,132	1,034

The unique method provided in the original design for handling grinding balls has been very successful. The balls are unloaded from railroad cars by magnet into overhead bins, from which they flow by gravity through counterbalanced gate spouts to a trailer drawn by a battery truck. Both are designed for rail travel. After being weighed on track platform scales, the truck and trailer travel along a 24-in. gauge railroad running on a platform between the mill-scoop housings and the sand-discharge ends of the classifiers. When delivered to the various grinding units the balls roll out of a side gate in the trailer into a rubber-lined chute, which diverts them into the drum-feeder section of the grinding-mill scoop. Two women with two trucks have been doing this work of charging 25 to 35 tons of balls in about 2 hr daily. Grinding balls are added to the mills only on day shift according to current average unit consumption multiplied by the mill tonnage treated per day. As there is no rapid method of determining the height of the ball charges in the shells, the mills must be shut down periodically every two to four weeks for inspections through the manhole. In most of the shells the ball levels are kept within 4 in. above or below the center line, which makes the weight of the ball charge vary between 50 and 60 tons.

The standard pattern of grinding-mill barrel liners has been the shiplap type. The original shape was altered by adding metal to the convex curve. The latest change has been the return to the original section for the half of the set nearest the discharge end, because of the smaller amount of wear occurring in that section of the mill. The maximum thickness of the barrel and feed-end liners is 5 in. Metal has been added to the feed-end liners to correct deficiencies revealed by wear curves.

Tests of two sets of rolled-steel-plate barrel liners, each 3 in. thick, with lifter clamp bars 2½ in. high above the liners, are in progress. Results after six months of operation indicate that liner life will be the criterion.

The grate assembly is composed of outer sectors having the bars in radial directions and inner sectors having the bars tangent to the various circumferences on which they lie. The bar sections are tapered outward from 1 in. to 11/16 in. in 3½-in. depth. Spaces between the inner edges of the bars of the outer sectors are ⅝ in. and between those of the inner sectors are ¾ in. Very little blinding of the grate openings by scrap balls occurs. The discharge end is equipped inside with 10 elevating sectors, which discharge the pulp through the trunnion. A helical splitter of local development is attached to the discharge trunnion for dividing the discharged pulp into equal portions to feed the two classifiers of each unit.

The first used feed-end liners were made in one piece, not including two manhole liners. At times the castings were warped, which deformed the mill head when the liner bolts were tightened. This caused weaving of the trunnion in the bearing, with resulting development of heat. The trouble was overcome by the manufacture of the feed-end liners in four pieces.

Average useful life of the principal liner parts of the grinding mills during the years 1943 to 1945, inclusive, was as shown in Table 7.

All wearing parts, except grates, from which the preceding figures were obtained were of the original lighter weight type. The weights per set of parts now being used are as follows: barrel liners, 46,680 lb; feed-end liners, 16,044; scoop lips (2), 514.

The consumption of manganese-steel liner parts of the mills averaged 0.15 lb per ton of ore ground during the four years 1942 to 1945, inclusive.

Separate lubrication systems, each containing about 2000 gal of oil, are provided for the mills of the original and extension plants. Electric heaters keep the oil temperature near 100°F. The oil is of the same type as is used in the crushers. Tests show that the average flow of oil to the trunnion and pinion bearings of each mill is 6.3 gal per minute. The oil of the two systems is cleaned by centrifuging on alternate days. One man is occupied continuously on that work except on every tenth day, when he operates the centrifuge in the primary crushing plant.

TABLE 7—*Life of Parts*

Equipment	Sets in Average	Life, Days	Weight per Set, Lb
Barrel liners, 48 per set.....	32	334	43,236
Feed-end liners.............	47	269	14,520
Grates and grate-clamp parts	35	271	10,127
Scoop lips, two per set......	28	53	386

Repair facilities for routine replacements are excellent. The grinding floor, which is almost 1000 ft long, has repair bays at each end and at the center. Three cranes service the grinding section. A 10-ton and a 25-ton crane work over the end bays or assist in light work in the grinding section. A double-hook 150-ton capacity crane is used to transport the grinding-mill shells to and from the repair bays.

Four spare mill shells and one spare discharge end are available for reconditioning in the repair bays. The loss of production time is no greater than two to three hours to change shells in the operating stands. Trunnion bearings are left on a shell when it is transferred. Repair stands that are appropriate for the work to be done are provided. Either the feed-end or discharge-end trunnion is supported in a bearing stand and the other end in a greased wooden saddle. Both ends of the shells rest in bearing stands when stored. End-liner or grate replacements are made while the head rests upon the end of the trunnion on the floor. Barrel-liner replacements require removal of the feed end head. The liner bolts are removed and the scrap liners drop free as the shell is revolved in the repair stand by a cable from the crane hook. New barrel liners are placed in the shells by the use of a battery-powered boom truck. Two boom trucks are available. The use of compressed-air impact wrenches makes the work of loosening or tightening liner bolts rapid and relatively easy.

The central area between the original and extension plants is used for repairs of the classifier spirals. In making classifier-shoe replacements, old bolts are cut off with an acetylene torch and new bolts are tightened with an air-powered angle wrench. For greater efficiency in reconditioning, the spirals are supported on bearings and revolved as repairs proceed. A lifting beam, which carries the spirals on the crane hook at the inclination on which they lie in the classifier, makes removal and installation easier. Four spare spirals are available for reconditioning. The amount of production time lost for replacement of a spiral is about one hour.

In addition to the expected regular replacement of wearing shoes and of a few helical flights of the spirals, some breakage of the 30-ft.-long shafts has occurred. Torsion and deflection effects caused failures after about two years of operation. The first repair method consisted of welding the broken ends of the shaft together over a sleeve pressed in the shaft. When correctly done that method gave about two years additional life. Adoption of a method being used at the New Cornelia Branch promises to be satisfactory in giving the rigidity required to resist breakage. Four bars, each 1 by 4 in. in cross section and extending to 3 ft from the ends of the spiral, are welded at 90° intervals to the arms that support the helical flights. The inner edges of the bars are 6½ in. from the shaft.

Flotation and Regrinding of Concentrate

Distribution of flotation feed that contains 18 pct solids is made by flow from outlets at the same level in the sides of a distributor that contains baffles to assist in mixing the returned cleaner tailing with the new feed (Fig 5). Collector reagent and lime having been added in the grinding process, frother reagents are dispersed in hot water and added to the flotation feed a short distance before it enters the distributor. Additional lime may be added at the same location if it is desired but that is done infrequently because addition to the grinding circuit has been found to be much more effective in controlling the alkalinity.

The mechanical flotation unit develops air for flotation and circulates the ore pulp for aeration by the action of concentric squirrel cages composed of rubber-covered posts. The inner or rotating squirrel cage has impellers at the top and bottom, which oppose each other by pumping into the rotor to force ore pulp out between the rotor posts. A vortex forms in the pulp by the pumping action of the upper impeller, into which air is drawn. The pulp is aerated as the mixture of air and pulp is forced through the openings between the posts of the stator. Probable action of the machine in the production of small bubbles is both by cavitation and by shearing.

The aeration units of the rougher cells are operated by 10-hp vertical motor reducers and those of the cleaner cells by 5-hp motor reducers. The rotor speeds of both rougher and cleaner rotors are either 410 or 422 rpm according to which manufacturer built the power unit. The motors are 1750 rpm induction type.

All parts of the aeration units that are subject to wear or corrosion are protected with rubber. A perforated rubber skirt surrounds the stationary squirrel cage at its top, which disperses the flow of pulp as it leaves the stator. The aeration units are set in long tanks, each having a single set of control gates, and the aeration units are removable singly for repair without the necessity of shutting down the entire machine.

The machines have met expectations in regard to ease of repair and the metallurgical results obtained by their use.

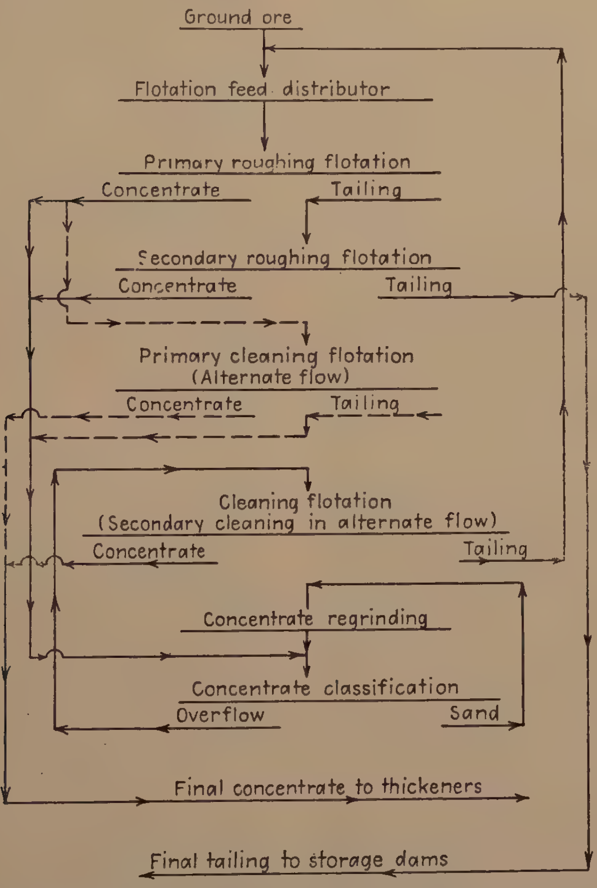
Ground ore
Flotation feed distributor
Primary roughing flotation
Concentrate
Tailing
Secondary roughing flotation
Concentrate
Tailing
Primary cleaning flotation
(Alternate flow)
Concentrate
Tailing
Cleaning flotation
(Secondary cleaning in alternate flow)
Concentrate
Tailing
Concentrate regrinding
Concentrate classification
Overflow
Sand
Final concentrate to thickeners
Final tailing to storage dams

Fig 5—Flowsheet of flotation and concentrate regrinding.
The feed distributor tanks are rectangular, 368 and 510 cu ft, respectively.
All flotation machines are mechanical level type, 144 cells in lines of four, 66 inch.
Secondary roughing is in 288 cells in lines of eight, 66 inch.
For primary cleaning there are 72 cells in lines of two or three, 56 inch.
Cleaning flotation employs 100 cells in lines of three and four, 56 inch.
Concentrate regrinding is in six 8½ by 12-ft overflow ball mills.
Classification is in six 54-in. duplex submerged-spiral classifiers, having four speeds.

Power consumption has been moderate and capacity good during the past four years.

The roughing flotation machines are arranged in rows of 12 cells. The first four are the primary rougher machines and 2 ft below them is the secondary rougher machine, containing eight cells. In the original plant there are two flotation-feed distributors, each feeding 10 rows of roughing cells divided into sections of five machines of 12 units each on two sides of the distributors. In the extension plant there are also four roughing flotation sections but each is composed of only four rows of machines.

TABLE 8—*Power Consumption, Flotation Unit*

Year	Kw-hr per Ton Ore	Tons Ore per Day per Sq Ft Area	
		Rougher	Total
1942	2.03	3.66	3.01
1943	2.11	3.71	3.19
1944	2.22	3.28	2.82
1945	2.37	3.38	3.07

One operator is employed for each flotation section, making a total of eight operators on roughing and two on cleaning flotation on each shift for capacity operation. They work under the direct supervision of two assistant shift bosses, who also supervise the work of the two operators in the concentrate-regrinding section.

Two samplers are employed on each shift to observe the operation of the automatic sampling systems, change sample containers and to take hand samples where necessary. A woman is employed on each shift to determine the copper contents of the ore, concentrate and tailing by routine analytical methods. Results are posted about an hour after the samples have been taken. The practice has been of much assistance in the operation of the flotation machines to produce the required grade of concentrate with the maximum extraction of copper.

All flotation sections are equipped with fluorescent lights, which minimize differences between day and night operation.

There are separate regrinding sections for the original and extension plants, each containing three 8½ by 12-ft overflow-type ball mills and three 54-in. duplex submerged-spiral classifiers driven by four-speed motors. The classifier spirals can be driven at 4.7, 3.1, 2.7 or 1.6 rpm but normally they are operated at either of the two lowest speeds, according to the mineral loads received by the classifiers. The regrinding mills are charged daily with an approximate 50 pct proportion of 2-in. chilled cast-iron balls and 50 pct scrap recovered from the ore-grinding circuit by the hydraulic ball traps.

Two employees on each shift operate the two sections and make periodic determinations of the alkalinity and the solid content of the classifier overflows. Sufficient lime is added to the regrinding mills to obtain the alkalinity required in the concentrate-cleaning operation.

The power required for regrinding includes the amount used for all pumping of the product to and from the cleaner flotation machines, and is 0.66 kw-hr per ton of ore treated by the concentrator.

First operation of the Morenci concentrator was done under a flotation flowsheet chosen during operation of the test concentrator. The primary rougher concentrate was re-treated in primary cleaning machines and the tailing from the latter was united with the secondary rougher concentrate, reground and refloated in secondary cleaning machines. The concentrate made in both cleaning operations composed the final concentrate. The use of that flowsheet, which is designated as alternate in the flowsheet shown in Fig 5, was abandoned as a result of later experiments to increase the grade of the concentrate. The cause was lack of sufficient control of the composition of the primary cleaner concentrate resulting from changes in the character of association of pyrite and chalcocite in the ore.

The flowsheet that is now in use was second choice in the test concentrator experiments because of the probable large regrinding capacity required, and because test results indicated that concentrate of a higher moisture content would be produced. The flowsheet includes regrinding of all rougher concentrate and single-stage flotation cleaning. It went into use in the Morenci concentrator in July 1945 after tests that determined that concentrate-regrinding capacity was sufficient and that the increase in the moisture content of the concentrate was minor because finer grinding could be confined mostly to the coarser sizes with but little increase in the proportion of the finest sizes. This was done by maintaining a relatively high solid content in the regrind-classifier overflow. The higher specific gravity of the pulp permits less of the finer sizes to settle into the spirals to be returned to the grinding mill. A chronological account of the various metallurgical changes made since the beginning of operation is given in the following paragraphs.

Metallurgical Changes

During February, March and April of 1942, after operation was begun, the concentrate contained only 17 pct copper. Little attention could be given to metallurgy because time was fully occupied in training personnel and in putting new equipment into service. During the remainder of the year the concentrate grade did not vary much from 21 pct copper. The result was obtained by decreasing the rate of froth removal from the flotation machines, by decreasing the number of primary cleaner flotation machines in use and by increasing the alkalinity of both primary and secondary concentrate-cleaning circuits. Those changes decreased the extraction of iron from 40 to 30 pct without greatly affecting the recovery of copper.

The use of a mixture of 25 pct of high-gravity fuel oil with cresylic acid was begun in October 1943 as a froth modifier, to produce a more voluminous and stable froth. The use of B-48 alcoholic froth modifier was abandoned because the free-flowing character of the froth had adverse effects upon the grade of concentrate.

During the year 1943 the copper content of the ore increased from about 0.85 pct in the first three months to 1 pct in the last six months. The iron content of the ore remained nearly constant, which made the ore a little more favorable to the production of a higher grade of concentrate. All attempts to increase the grade without decreasing the extraction of copper by making ordinary changes such as the variation of flotation-machine operating methods, and in quantities of reagents used, were unsuccessful. The amount of regrinding of concentrate that could be done was limited because any increase in moisture in the concentrate was critical in smelter operation at that time. The average copper content of the concentrate during 1943 was 21.38 pct without much change from that figure during individual months.

A continuation of the increase in the grade of ore without an increase in the iron content of the ore in the year 1944 made it more favorable toward the production of a higher grade of concentrate. The ratio of copper to iron in the ore decreased from 3.6 to 1 during the first three months to approximately 3 to 1 during the last six months. The grade of concentrate was increased from 21 pct copper during the first three months to more than 26 pct during the last six months of the year 1944. The smaller proportion of iron to copper in the ore made the changes in operating technique that were begun in the year 1943 more effective. Average extractions of copper and iron for the year were essentially unchanged from those of the preceding year.

Laboratory investigations during the year indicated that quebracho had some value as a reagent to assist in the rejection of pyrite either as a dispersant, as a reducing or neutralizing agent for soluble salts that activate pyrite, or as a direct depressant of pyrite. Other reagents having similar effects were disclosed. The laboratory tests were followed by a plant test in the concentrate-cleaning section of a part of the concentrator beginning in November 1944.

The proportion of copper in the ore increased in 1945 to 1.15 pct. That increase, with the proportion of iron remaining constant at 3.5 pct, kept the ratio of iron to copper in it at 3.0 to 1. An increase in the proportion of oxidized copper also occurred as a result of mining some higher grade ore adjacent to old stoping areas. The low iron to copper ratio was favorable to the production of a high grade of concentrate but the activating effects on pyrite of the soluble salts accompanying the larger proportion of oxidized copper was an adverse influence. The net result was that the grade of concentrate made during the last half of 1944 was continued through 1945 with an average copper content of 26.81 pct, and without further increase in the loss of sulphide copper in the tailing.

The test of quebracho in the process for re-treatment of concentrate in part of the concentrator was continued through May 1945. Beginning in June the use of that reagent, or No. 610 reagent, which is a lignin product having similar effects, was used in the entire concentrator.

During July 1945 a test run was made in part of the concentrator in which refloating or cleaning the primary rougher concentrate was discontinued and all rougher concentrate was classified and ground. The moisture content of the final concentrate was not increased greatly, so later in the month the flowsheet of the entire concentrator was changed to eliminate primary cleaning. It is probable that the dispersing reagent assisted in eliminating gangue slimes from the concentrate, which might have caused an increase in the moisture because the insoluble content of the concentrate decreased from 5.9 to 5.1 pct following the universal use of quebracho or No. 610 reagent. In the first month of the universal use of the reagent that antedated concentrate regrinding experiments, the insoluble in the concentrate decreased to 4.8 pct from 5.6 pct in the preceding month.

Effects of the flowsheet change upon the concentrate screen analysis and moisture content are shown in Table 9.

TABLE 9—*Effects of Change in Flowsheet*

Year 1945	Concentrate Operation	Final Concentrate		
		Screen Analysis, Pct		Moisture, Pct
		On 200 Mesh	Through 325 Mesh	
Jan. to June...	Primary cleaning	19.8	65.3	9.39
July to Dec...	No primary cleaning	10.3	70.1	9.65

A 48 pct decrease resulted in the amount of sizes on 200 mesh while the finest sizes increased only by 7.3 pct. The moisture content of the concentrate increased only 2.8 pct.

The chemical composition of the concentrate produced in the year 1945 is typical of present practice: copper, 26.81 pct; insoluble, 5.5; silica, 3.8; alumina, 1.4; iron, 27.9; lime, 0.1; sulphur, 38.7; molybdenum sulphide, 0.24; gold, 0.019 oz per ton; silver, 1.109 oz per ton.

The calculated pyrite content of this concentrate is 59.9 pct. The molybdenum sulphide content has increased from 0.13 pct in 1942. No effort is made to recover molybdenum. Arsenic, antimony, lead, nickel and cobalt are present in minute amounts.

The recovery of the precious metals is incidental. Gold extraction in the concentrate in the year 1945 was 55.65 pct of the amount in the ore and silver extraction was 77.30 pct.

The principal reagents used during the four years of plant operation are listed in Table 10.

The increase in the amount of frothers used is the result of the progressive decrease in the amount of altered ore received for treatment as the development of the ore body progressed.

No. 208 collector reagent is added to the grinding-mill feed as a 5 pct water solution. Bucket-wheel feeders are used to control the rate of addition of all reagents except lime.

TABLE 10—*Reagents Used*
POUNDS PER TON ORE CONCENTRATED

Year	1942	1943	1944	1945
Lime	3.327	3.359	3.500	3.624
Collector Reagent No. 208	0.033	0.037	0.033	0.031
Frother reagents:				
Cresylic acid No. 4020	0.065	0.064	0.093	0.098
Fuel oil, 11 Bé	0.005	0.017	0.020	0.016
No. B-48 alcohol	0.008	0.001	0.001	
Pine oil	0.001			
Total frothers	0.079	0.082	0.114	0.114

The frother reagent mixture of 86 pct cresylic acid and 14 pct fuel oil by weight is dispersed in hot water and added to the flotation feed. The device for effecting dispersion was designed locally and consists of a small cylindrical tank in which there is a horizontal disk and a pipe at 90° to it, with a small orifice in the end. A 3450-rpm centrifugal pump circulates the mixture of reagents and hot water that disperses the reagents by the impact of the submerged jet against the disk.

Quebracho was added to the feed of the concentrate-regrinding circuit as a 5 pct solution in water at the rate of 0.10 lb per ton of concentrate treated. It had to be dissolved in hot water and left a gummy residue in the bottom of the dissolving tank, which was periodically removed by dissolving in additional hot water. It also contains a fibrous component, which choked feeders and pipe lines and made the rate of addition erratic.

The lignin compound, No. 610, is easily soluble in cold water and leaves no residue. The cost is about half that of quebracho. It is dissolved as a 5 pct solution and added at the rate of 0.10 lb per ton of concentrate treated.

At present burnt lime is purchased from a plant near Douglas, Arizona, but early construction of a rotary-kiln plant on the site of the reduction works is planned. The lime now being used contains 80 to 85 pct available CaO. It is hydrated to a milk containing 10 pct solids in a plant consisting of a 7½ by 12-ft ball mill and a 60-in. high weir duplex spiral classifier operating in closed circuit with the ball mill. The milk of lime is pumped to agitator storage tanks having a total capacity of 500,000 gal. It is fed from the storage tanks by float valves to two circulation-feed tanks, from which it is pumped through and returned from all parts of the concentrator where lime is used. Most of the lime is added to the grinding mills, the rate being regulated by diamond port iron plug valves equipped with dials. Auxiliary outlets are provided at each mill for greater rates of addition when oxidized ore is treated.

Alkalinities being maintained at present range from 3 to 5 grains CaO per gallon of water with the pulp in the roughing flotation machines to 10 to 15 grains CaO per gallon in the concentrate-cleaning flotation operation. Some freedom of adjustment is granted to the operators in order to secure the desired type of froth in each operation.

A reagent-storage and preparation building has been provided in which all reagents are dissolved and pumped to storage tanks in the concentrator building. The lime-hydrate circulation tanks and pumps are also in this building.

FILTRATION OF CONCENTRATE

The filtration flowsheet is shown in Fig 6. Concentrate pulp containing about 20 pct solids is laundered to the thickeners and after thickening to 55 or 60 pct solids is elevated and flows in launders to the filter plants. There are two filter plants, one containing four filters and three rotary vacuum pumps and the other containing three filters and two rotary vacuum pumps.

Two series of belt conveyors deliver the filtered concentrate to the storage beds. Either filter plant can deliver concentrate to either of the two main conveyors by moving a connecting conveyor.

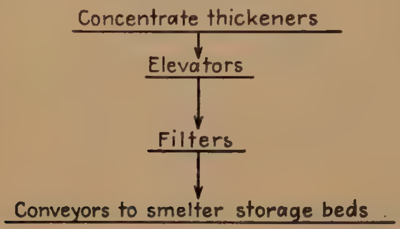

FIG 6—FLOWSHEET FOR FILTRATION OF CONCENTRATE.

There are five torque-type thickeners of 100-ft diameter; six 30-in. belt and bucket elevators in three towers; seven seven-disk filters 8½-ft diameter; and two 20-in. belt conveyors.

There are two small rotary positive-pressure blowers in each filter plant, which provide air at one pound per square inch pressure for the discharge of the filter cake.

The rotary vacuum pumps have a rated capacity of 1875 cfm at 26 in. of mercury vacuum. They are direct-driven at 495 rpm by 100-hp motors. The pumps are protected from the entrance of concentrate by air cleaners and a barometric seal on the inlet pipe line. The vacuum is kept near 21 in. of mercury by hand regulation of an air-inlet valve on the pipe line. The vacuum formerly was kept at 23 in. of mercury but no drier filter cake was made than could be obtained at 21 in. vacuum, and the pumps operated with a lower exhaust temperature. When 21 in. of vacuum is produced, the temperature of the exhaust varies between 400° and 450°F. Rotors and cases are cleaned yearly and at that time it is usually necessary to replace rotor and floating ring blades.

Sirens operated by mercury-switch water-level gauges warn the operators of excess filtrate in the tanks. Filtrate is pumped to the concentrate thickeners and other pumps return filter overflow or drainage to the thickeners.

Standard sector covers made of cotton filter fabric are used. During the earlier years of operation, holes in the sector covers were patched with filter fabric attached with waterproof cement. During the year 1945 the practice was confined to the patching of only covers that had not been in use long enough for the pores to be blinded with lime scale. This change in practice was made in order to maintain filter capacity and the lowest possible moisture in the concentrate. During that year the sector covers had an operating life of 102.5 days.

The discharge from disk filters is periodic and irregular in amount. This has caused difficulties in weighing it by conveyor scales because of the variable load upon the belt. Regulation of the discharge onto the weighing belt has been solved by a method that requires the installation of an auxiliary conveyor under each filter, having the speed synchronized so that the belt travels the length of the filter in the same time that one of the 10 rows of sectors travels through an arc of 36°. This places an empty section of conveyor under the filter each time that a row of sectors is discharged. The installation of one of these auxiliary conveyors has been completed and operated with the desired results. Six additional conveyors are to be installed.

Various tests have been made to reduce the moisture content of the filter cake but without any great success because of the relatively low moisture content. The tests included heating of the filter feed, the use of wetting agents, a trial of synthetic-fabric sector covers and several combinations of disk speed and cake-building time. A tangible factor that has affected the moisture content of the concentrate has been its screen analysis. Limits to which grinding of Morenci concentrate may be carried without causing a large increase in moisture content are approximately 10 pct on 200-mesh, or 70 pct through 325-mesh. There are intangible factors such as slimes from altered ore and flocculation of the filter feed. The effects of those factors have not been definitely related to the concentrate moisture content.

Two operators and two helpers on each shift are required for concentrate thickening, filtration and delivery to the conveyors that transport it to the smelter. Operating results for the past four years are tabulated in Table 11.

TABLE 11—*Filtration Data*

Year	1942	1943	1944	1945
Dry tons product per filter day	368.6	399.1	438.9	357.5
Moisture in product, pct	9.30	9.15	9.43	9.52
Power per ton ore milled, kw-hr	0.15	0.13	0.18	0.18
Power per ton concentrate handled, kw-hr	4.24	3.46	4.47	4.83

About 6 pct of the power figure is the amount used for conveying the concentrate to the storage beds.

WATER SUPPLY AND CONSUMPTION

Water supply and consumption is given in detail in Fig 7. The source of the fresh water supply is Eagle Creek, from which it is pumped a distance of 6 miles against a difference in elevation of 1100 ft. The water has low total solid and chlorine contents, as shown in the typical chemical analysis of the composite sample taken in July 1945 (Table 12).

TABLE 12—*Chemical Analysis of Water, July* 1945
PARTS PER MILLION

SiO_2	21.0	HCO_3	250.0
Fe	None	CO_3	3.0
Al	None	Organic and volatile	72.0
Ca	41.1	Solids by analysis	535.7
Mg	14.0	Residue	351.0
Na and K	63.4	Alkalinity	210.0
Cl	5.14		
SO_4	19.8		

The determinations were made by the scheme of analysis used by the American Public Health Association.

Consumption of fresh water has averaged 213 gal per ton of ore milled during the first four years of operation and was near that figure for each year, but during individual months it varied considerably because of changes in the amount of water sent to and reclaimed from the tailing dams. The settling rate of the ore governs the amount sent to the dams, while weather conditions and peculiarities of the storage dams upon which the tailing is being disposed determine the amount possible to return. Fresh-water consumption varied during the year 1945 from a low of 168 gal per ton of ore to a maximum of 267 gal and averaged 218 gal. Consumption of both fresh and reclaimed water is about five times that of fresh water, or 50 million gallons per day. Of that amount about 42 million gallons is circulated by the water-pumping installations of the concentrator at the average rate of 29,000 gal per minute.

The solid content of the flotation tailing or tailing thickener feed is about 18 pct and the underflow is discharged at 44 to 47 pct solids. As originally designed, the tailing-thickening installation provided 519,000 sq ft of settling area, or 11½ sq ft per ton of tailing to be thickened. It has since become necessary to convert one 325-ft-diam thickener for use as auxiliary fresh-water storage, which leaves 436,000 sq ft of area, or 9.6 sq ft per ton of tailing. That unit area is sufficient for most settling requirements but there have been times when the solid content of the underflows had to be decreased in order to prevent overflow of slimes.

There are three underflow pipes, each 8 in. in diameter, from each tailing thickener. Pipes from five thickeners converge through a system of tunnels to one discharge or spigot house, and those from the two remaining thickeners end in another house. There are two main-line diaphragm valves on each line under the centers of the thickeners. The underflow lines are cement-asbestos composition pipe. No replacements of the pipe have been made as a result of wear, but as plug valves are used near the discharge end to cut off flow when changing orifices they must be closed and opened slowly to avoid breakage by water hammer. Following the plug valves, the 8-in. pipe is joined to a steel pipe 4 in. in diameter, on the end of which is a cam lever clamp for holding the orifices in place. Orifices are made of conveyor-belt rubber containing several layers of fabric.

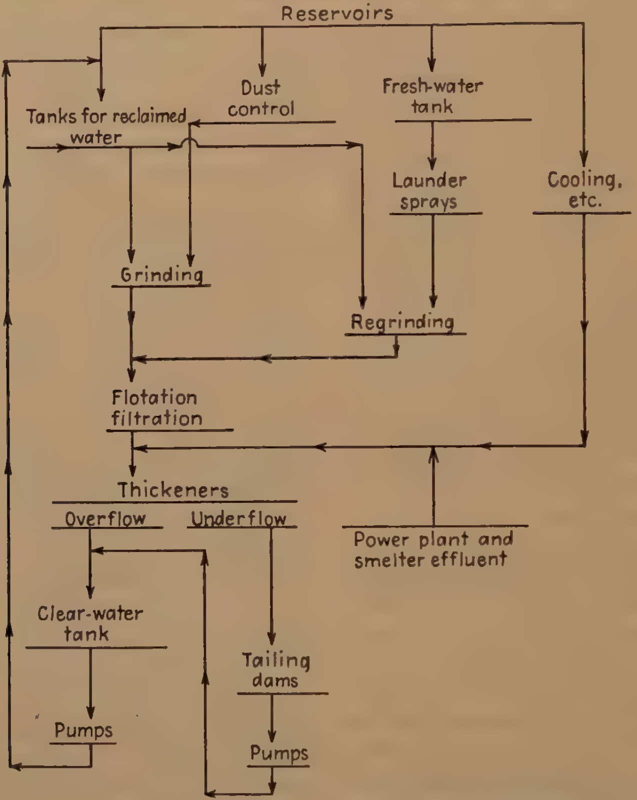

FIG 7—FLOWSHEET OF WATER SUPPLY AND CONSUMPTION.

The reservoirs hold 17 million gallons; three reclaimed-water tanks, 100,000 gal each; one fresh-water tank, 1,000,000 gal.

Tailing thickeners, traction, have dual drive. Five 300-ft diam; one 325-ft diam. (One has been converted to a water reservoir.) Clear-water surge tank holds one million gallons.

There are two concentrator pumping stations, equipped with ten 5000-gpm pumps.

There is one feeder pumping station with four 1000-gpm pumps; and one feeder station with seven 1000-gpm pumps. One main station has seven 1000-gpm pumps and one 50,000-gal tank.

Reclamation of water from the tailing dams being an integral part of tailing disposal at Morenci, it is described under that heading in the following section of the paper.

One pump operator and one thickener operator are required on each of the three shifts for thickening tailing and pumping reclaimed water.

The power consumed in pumping reclaimed water in the years 1944 and 1945 after operations were stabilized amounted to 1.53 kw-hr per ton of ore treated, or 10.4 pct of the total power requirements for crushing and concentrating. The figures quoted include the power consumed in returning water from the tailing dams, which in July 1945 was 20.9 pct of the total power required for all pumping of reclaimed water.

Tailing Disposal

The tailing is stored in a wide valley in the foothills south of Morenci (Fig 8). The bottom of the valley is divided into three ravines, the central and widest of which is known as Morenci Canyon. To the west of Morenci Canyon is Stargo Canyon, separated from it by low ridges. Along the east side of Morenci Canyon is a comparatively high ridge traversing the entire length of the tailing-disposal area. A narrow ravine to the east of the ridge has been named East Canyon. All three of the ravines lie on an average slope of 4 pct. The upper parts of Morenci and East Canyons were used for disposal of tailing from the former No. 6 concentrator. The whole area is roughly 3 miles north-south and 1 mile east-west in size. The difference in elevation between the upper part nearest the reduction works and the lowest near the reclaimed-water pumping stations is approximately 700 ft.

Tailing flows to the disposal area through a flume constructed on the ridge between Morenci and East Canyons. The flume is 17,000 ft long and is built on a minimum grade of 1.5 pct and a maximum of 2.5 pct. The average grade is 1.75 pct. The original flume was a curved-bottom launder made of cement-asbestos composition material. Because of wear along the center line of the bottom, the flume is now being sheathed in concrete. The curved bottom was designed to carry small flows without impedance.

Gates are located at appropriate points along the flume by which the tailing can be diverted to the various storage dams. Laterals from the flume to the dams are concrete launders having inside sectional dimensions of 36 in. wide by 24 in. deep, with walls 6 in. and base 8 in. thick. Smaller dams, which will not accommodate full flow of tailing, have feed launders of 30 by 20 in. inside dimensions. Immediately preceding their junction with the distribution pipes on the dams, the laterals are built with a cover extending for a minimum of 5 ft in elevation for every 1000 ft of distribution pipe, to provide a section in which pressure is developed for forcing the tailing through the level pipe for distribution on the dams.

The distribution pipe is 24 in. in diameter. It is made in 13-ft sections of cement-asbestos composition, to withstand a working pressure of 100 psi. It is supported upon wooden trestles 40 ft high when first constructed. The trestles have walkways with railings and toe boards. Valve-operating platforms are built on the opposite side of the pipe from the walkways. All trestles are equipped with incandescent lighting. Tailing is discharged toward the bank through 6-in. lubricated plug valves 52 ft apart on the pipe. Experiments have indicated superior washing of slimes toward the back of the dam or classification of sand near the front of the dam by substituting 8-in. valves for the 6-in. valves. Ten-inch valves are also to be tried. When a trestle is new and farthest from the bank, the valves are discharged into V-section troughs about 10 ft long, set on a slope of about 20° from the horizontal to drop the stream closer to the bank.

Banks are built 30 or 36 in. high and each bank is set in toward the trestle, so that the face of the dam has a slope of 1 ft of vertical height in 2 ft of horizontal distance. They are constructed by several methods, the choice of which is dependent upon the dryness of the dam when the bank is built.

If the deposited tailing is still soft, the bank on that dam is built of old dry tailing produced during operation of the former

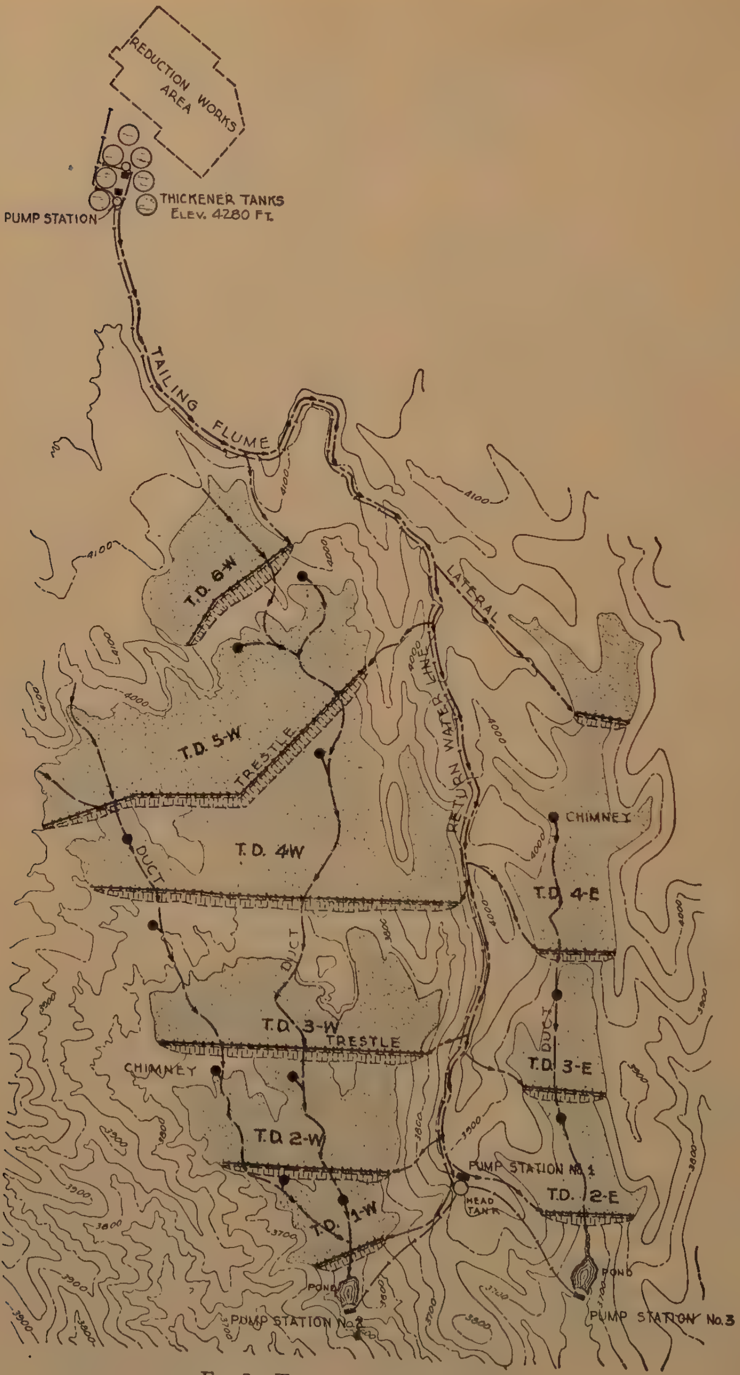

Fig 8—Tailing-disposal area.

No. 6 concentrator. The material is loaded by a small power shovel and transported in dump trucks to the dam. This type of bank is built to a height of 30 in. in two lifts or layers, each of which is spread, compacted and leveled by the use of small wide-tread angledozers. Greater height than 30 in. would require the handling of a disproportionate quantity of dry tailing, as the top must be wide enough for truck travel.

If the deposited tailing is dry enough to support a dragline excavator on timber mats, the new bank is built by that method using the deposited tailing. The bank is compacted to a height of 36 in. by an angledozer. A maximum height of 36 in. was chosen for this kind of bank, to avoid the formation of too deep a borrow pit adjacent to the bank. That height has also been found to produce more rapid drying of the deposited tailing than a deeper deposit, because of the limited absorption of moisture by the previously deposited tailing. Divisions are left in the borrow pit at about 200-ft intervals, to decrease the trapping of slimes near the bank. The more rapid overflow from each section washes the slimes toward the pond. If the deposited tailing of which the bank is made is sufficiently dry, the borrow pit is partly refilled by the use of bulldozers, or the pit depth is kept at a minimum by transporting the tailing to the dragline with bulldozers.

If the drying period of the deposited tailing is long enough, the most desirable type of bank is that built by the use of bulldozers. Two of the machines are used, one to bring material toward the bank and the other to elevate and compact it. This method of bank building produces well compacted bank with a wide, shallow borrow pit behind it.

The earth-moving equipment is used also for construction of heavy dirt mats, upon which new trestle is constructed when insufficient drying time can be allowed for consolidation of the tailing.

The total area of the tailing-disposal site is about 3.15 sq miles, or 2000 acres. The useful storage area on it was approximately 533 acres at the end of the year 1945. That is 11.8 acres per 1000 tons of tailing to be stored daily. Two minor dams of a total area of 29 acres have since been filled to their maximum capacity, but additional area developed on the remaining dams has kept the useful storage area unchanged.

There are six dams in Morenci and Stargo Canyons, approximately 1500 ft from front to back. At the beginning of development four of the dams in the two canyons were in line but were separated by the low ridge. At the present time three of the four have become single dams by submergence of the ridge in tailing.

There were four dams in the East Canyon, but the one at the highest elevation has been filled to capacity. The one next below it is about 3000 ft long and is filled from both front and back to control the pond location for maximum reclamation of water.

Except when distribution of tailing is scattered among several dams, only one operator and three helpers are required for tailing disposal. One 6-in. valve will discharge about 2000 tons of tailing daily, so in order to dispose of 45,000 tons of tailing daily 22 valves must be opened. As they are 52 ft apart, it requires the care of 1200 ft of bank. The operator and a helper attend to that while the other two helpers drain water into the ducts or attend portable pumps if they are being used. They work without any other supervision than that of the operator, who is lead man on the shift and is responsible for correct performance of all required duties.

Facilities provided for reclamation of water from the dams consist of three concrete ducts extending under all of the dams in Stargo, Morenci and East Canyons. The two former converge at one settling pond, preceding a pumping station, and the duct of East Canyon ends at another. Sections of the ducts were built as needed early in the development of the dams and were connected later. In interior dimensions they are 32 in. wide and 34 in. high, with walls 6 in. thick and a minimum thickness of top and bottom of 8 in. Holes 8 in. in diameter, in the walls at a difference in elevation of 2 in. between holes, are used for withdrawing clear water from the ponds that form back of the dams.

The ducts were constructed on undisturbed ground and generally they follow the original grade of the canyons. Approximate lengths of the three ducts, including branches, are as follows: Stargo Canyon, 11,000 lin. ft; Morenci Canyon, 13,000; East Canyon, 6,000; total, 30,000.

The Stargo and Morenci Canyon ducts are united into one duct of the same sectional area under the lowest dam. The mouth is equipped with a large trash-screen compartment.

As the ducts eventually would become covered with tailing, concrete stacks were constructed with connections to the ducts. They have 10-in. diameter holes on a spiral line around the wall at a difference in elevation of 2 in. between holes. There are nine holes for each turn of the helix. The stacks have an inside diameter of 6 ft and walls 12 in. thick. Several of the stacks are now in use, since some of the ducts between dams have been covered with tailing. Holes in the ducts and stacks are closed with tapered wooden plugs when the tailing level reaches them.

Some of the ponds form in locations that are isolated from the ducts. In order to reclaim the water from them portable pumping units are used. They are built upon I-beam skids using automobile engines and open impeller pumps of 4 and 6-in. sizes. Flexible suction hoses equipped with check valves and skimming pipes are supported on floats made of oil drums. Light-weight portable pipe that can be easily handled and rapidly assembled conveys the water from the pumps to the ducts.

The settling ponds at the duct mouths are kept about half full of water, to allow room for a sudden influx of water. Spillways are provided to allow flood water to overflow to safety dams below the pumping stations. Water is pumped from the safety dams to the settling ponds as soon as possible.

No. 2 pumping station is below the settling pond at the junction of Morenci and Stargo Canyons. No. 3 pump station is below the settling pond in the East Canyon. Both deliver to a 50,000-gal tank at No. 1 pumping station on the lower end of the ridge between Morenci and East Canyons. There are seven single-stage 1000-gpm pumps in No. 2 station, which elevate the water 305 ft. No. 3 station contains four single-stage 1000-gpm pumps, which operate against a difference of elevation of 227 ft. No. 1 station is equipped with seven two-stage 1000-gpm pumps, which return the water to the concentrator pump stations through a 14-in. pipe line, part steel and part cement-asbestos, and a 16-in. steel pipe line against a static head of 425 ft. The pumps of No. 1 station operate by automatic controls actuated by floats in the feed tank. Two men on each shift operate all the pumps. Dial-operated telephones connect all pump stations with those at the concentrator.

The amount of water reclaimed from the dams varies from 1,000,000 to 6,000,000 gal daily and averaged 3,100,000 gal per operating day during 1945. That amount is equivalent to 91 gal per dry ton of ore treated in the concentrator during the year.

ACKNOWLEDGMENTS

The author thanks the management of the Phelps Dodge Corporation for permission to publish this account of the first years of operation of the Morenci concentrator.

Mufulira Copper Mines Limited, Concentrator, Northern Rhodesia

By Jack White* and Ralph B. Adair,† Members AIME

(New York Meeting, February 1948)

The Mufulira mine in Northern Rhodesia is 13° south of the Equator and at an altitude of 4100 ft above sea level.

The concentrator was planned in 1930 to treat about 10,000 tons of ore per day, but before construction work had gone very far the size of the plant was reduced to 1500 tons daily capacity. Then just as the construction was finished in December 1931, the price of copper did not warrant the commencement of operations and it was not until October 1933 that the mine re-opened.

Extensions have been made to the plant in 1934, 1935, 1936, 1937, 1940 and finally in 1945. The capacity of the plant when the 1945 extension is completed will be 9500 short tons per day. The various extensions have been carried out with the minimum confusion and with great skill on the part of the designers and the erection engineers.

In dealing with the plant in this paper it is proposed to give an outline of the present practice and only to refer to past practice when it is considered of particular interest.

Ore Treated

The ore consists of fine grained highly siliceous, felspathic and argillaceous quartzites, the latter containing graphitic carbon and locally termed graywacke. These rocks are considerably altered near the surface and on the fringes of the ore body. Despite extensive efforts at mixing, both underground and in the crushing plant, many sudden changes of ore take place in the mill feed.

The minerals are chalcocite, bornite and chalcopyrite with relatively unimportant amounts of malachite, native copper, azurite, covellite and chrysocolla. Gold is present at about 1 oz to a thousand tons, and silver at about 2 dwt. per ton. With increasing depth the nonsulphide content or, as commonly termed locally, the oxide content of the ore is decreasing and is now less than 0.25 pct against a high of 0.77 pct.

Crushing Operations

The first stage of crushing is done underground where the oversize from 12-in. grizzlies is fed to a 56 × 72-in. jaw crusher.

At the Selkirk shaft two 10-ton skips deliver the run of mine ore at a rate of 450 tons per hour to the 400-ton capacity shaft bins. One 6 ft wide Ross chain feeder and one locally made air-operated finger feeder feed the two 30-in. McCully crushers. Ahead of each crusher is a 6 × 9-ft spring grizzley spaced at 5 in. on a 35° slope. Neither the Ross chain feeder nor the finger feeder is entirely satisfactory and for the wet, muddy and sticky ore that has to be handled it is planned to replace these with two 6-ft wide apron pan feeders.

The McCully crushers deliver a minus 5-in. product, at a rate of over 500 tons per

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hour, which is conveyed by a series of 48 and 42-in. wide conveyor belts to four 600-ton capacity storage bins which serve as surge bins to four 5½-ft Symons Standard cone crushers.

Feeders from these bins are of the drum type on to 28-in. wide conveyors. At the head of each conveyor is a 43-in. diam magnet for the removal of tramp iron. One double deck 48 × 102-in. Gyrex screen ahead of each Symons helps keep the crusher feed free from mud. The upper deck of the Gyrex screens is fitted with ¾-in. thick cast manganese steel screens with 2½-in. square holes. The lower deck has 0.46 × 0.75-in. slotted wire screens. The undersize from these screens, which is usually very muddy, goes straight to the fine ore bins without any further crushing. The Symons product with the crushers set at about 1½ in. is conveyed by means of a scissors conveyor to 5 surge bins each of 400 tons capacity. Each bin delivers ore to two double deck Gyrex screens (60 × 102 in.) through roll type feeders. The upper deck of these screens is 1-in. square screen cloth and the lower 0.396 × 0.75 in. slotted screen. The oversize from both decks of two Gyrex screens feeds one of the 5½ ft Symons Shorthead crushers. 43-in. diam magnets at the discharge point of each belt feeding the Shortheads protect them, with their setting of about ¼ in., from tramp iron that may have passed the magnets on the Standard Symons crushers. Undersize from the lower deck joins the undersize from 48 × 102-in. Gyrex screens as a finished product from the crushing plant. This is taken by a 42 in. conveyor over a Merrick Weightometer to the fine ore bins. The discharge from the five shorthead crushers joins the product from the Standard Symons and is conveyed via the scissors conveyor back to the surge bins ahead of the Shorthead crushers. This makes a closed circuit, the only outlet being the undersize of the lower deck of the ten 60 × 102-in. Gyrex screens.

Oiling of all nine Symons crushers is done from a central pumping station. Each crusher has a sight feed and the whole system is interlocked with pressure switches which automatically trip the crushers if the pressure of the oil falls below a safe point.

Electrical interlocks are installed on all conveyors feeding crushers and on conveyors taking the discharges so that if any vital conveyor trips or is stopped, all the flow of ore leading to it is simultaneously stopped.

The sticky nature of the ore, which averages over 5 pct moisture, necessitated careful design of transfer chutes and even so, chokes at these points occur from time to time.

The life of crusher wearing parts is given in Table 1.

TABLE 1—*Life of Crusher Wearing Parts*

Part	Average	
	Hours Run	Tons Treated
30-in. McCully breaking heads....	1,800	950,000
30-in. McCully lower concaves....	1,240	650,000
5½-ft Standard Symons mantles..	3,890	479,000
5½-ft Standard Symons bowl liners	3,270	400,000
5½-ft Shorthead mantles.........	1,700	170,000
5½-ft Shorthead bowl liners.......	1,700	170,000
10-ft × 72-in. Hardinge inlet liners	4,365	150,947
10-ft × 72-in. Hardinge wrapper liner........................	3,597	123,554
10-ft × 72-in. Hardinge discharge liner........................	4,405	152,525

A comfortable capacity of the entire crushing plant is 500 tons per hour, and a typical screen analysis of the product is shown in Table 2.

The crushing plant flowsheet is given in Fig 1.

FINE GRINDING AND CLASSIFICATION

The fine ore bins have a live capacity of some 9000 tons (750 tons per mill). Ore is drawn from them by three roll-type feeders for each mill driven by variable speed reducers. (The last three mills installed employ vibrator-type feeders.) A separate

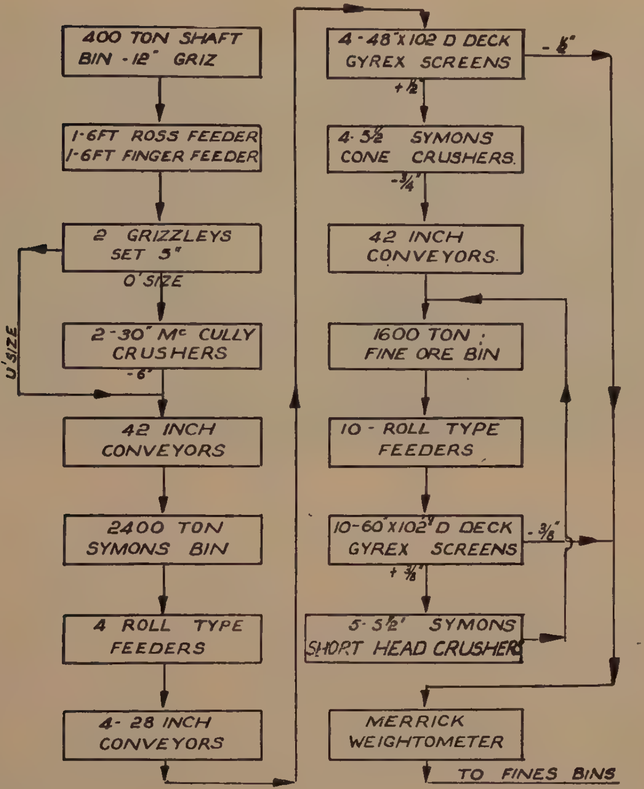

FIG 1—FLOWSHEET OF ORE-CRUSHING PLANT AS OF JANUARY 1, 1946.

Merrick Weightometer is fitted to each of the 12 ball-mill feed conveyors.

TABLE 2—*Typical Screen Analysis of Crushing Plant*

Screen Size	Weight, Pct	Cumulative Weight, Pct
+ 0.742 in.	0.6	0.6
+ 0.525 in.	3.4	4.0
+ 0.371 in.	13.4	17.4
+ 3 mesh	15.9	33.3
+ 4 mesh	8.8	42.1
+ 6 mesh	8.0	50.1
+ 8 mesh	7.5	57.6
+ 10 mesh	4.8	62.4
+ 14 mesh	4.3	66.7
+ 20 mesh	3.4	70.1
+ 28 mesh	3.2	73.3
+ 35 mesh	3.0	76.3
+ 48 mesh	2.5	78.8
+ 65 mesh	2.6	81.4
+100 mesh	2.7	84.1
+150 mesh	2.5	86.6
+200 mesh	2.4	89.0
+325 mesh	2.1	91.1
−325 mesh	8.9	

There are 11 open-end Hardinge ball mills (10 ft × 72 in.) and one No. 99 Marcy grate discharge ball mill. Grinding is single stage and the final product overflows an 8 × 30 ft Dorr DSFX classifier in closed circuit with each mill. The grind is between 58 and 60 pct minus-200 mesh and each mill handles about 34.5 tons per hour. The grinding circuit is extremely

TABLE 3—*Typical Screen Analysis of Classifier Overflow*

Mesh	Weight, Pct	Cumulative Weight, Pct
+ 65	4.45	4.45
+100	11.75	16.20
+150	17.05	33.25
+200	13.15	46.40
+325	12.90	59.30
−325	40.70	

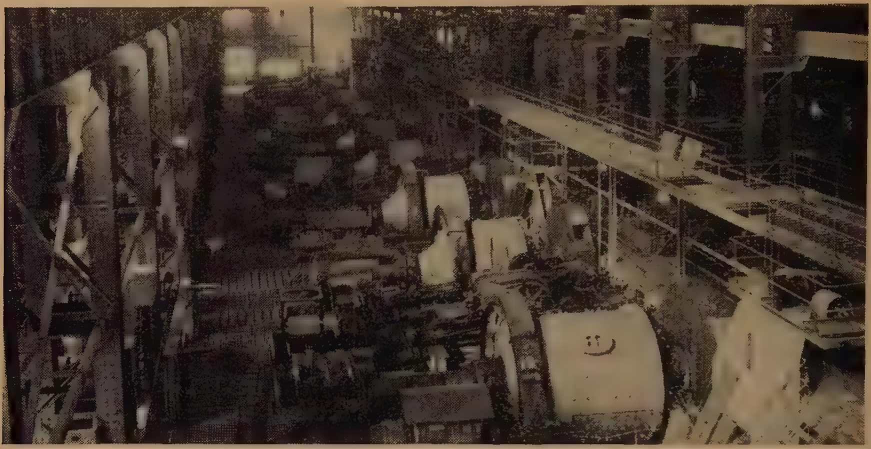

FIG 2—BALL-MILL BAY.

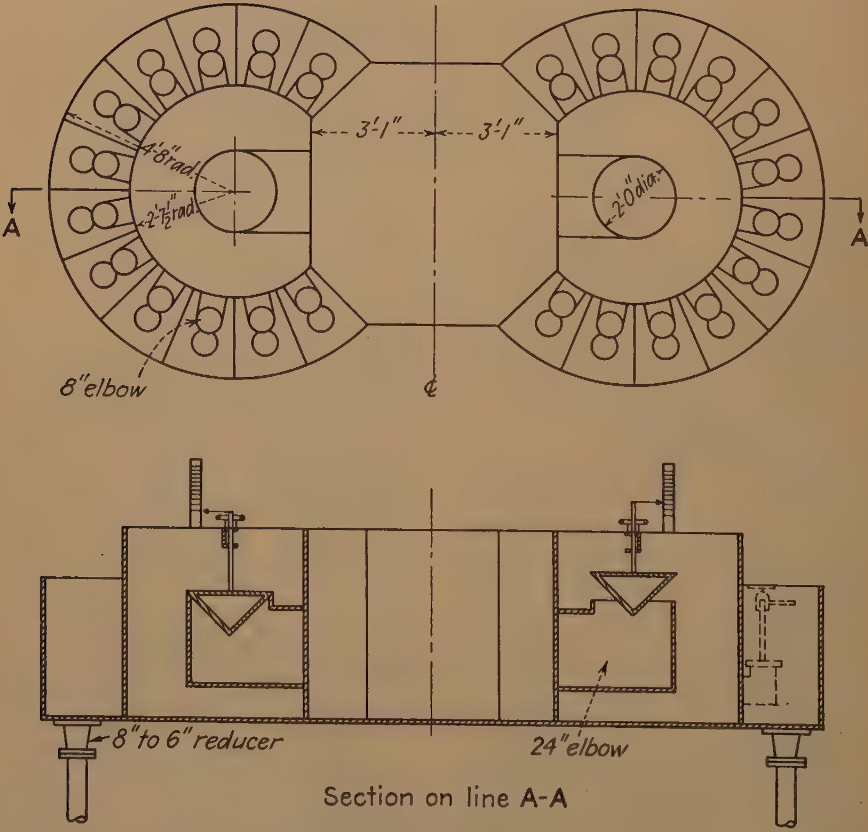

FIG 3—FLOTATION DISTRIBUTOR.

simple, with only one stage of classification. No heavy pumping costs are involved. A typical screen analysis of the classifier overflow is given in Table 3.

Circulating loads are held at about 400 pct and the classifier-overflow density is 23 pct solids by weight.

Between each ball mill and classifier the pulp passes over a modified Denver mineral jig. Until recently there was enough fine native copper to warrant the operation of the jigs, as some of the fine native copper was being lost in the tails. The product from all jigs is tabled on one Wilfley table and the concentrate from this table assays over 60 pct copper and about 2 dwt. of gold and 40 dwt. of silver. This concentrate is stockpiled and cleaned over another table at intervals, and a high-grade copper-gold concentrate is made which assays over 90 pct copper and between 5 and 10 oz of gold per ton.

The decreased amount of native copper now present in the ore makes the value of the jigs doubtful, and their operation may be discontinued in the near future.

After experimenting with different size balls (from 5 to 2 in.) 3-in. diameter was chosen as the best, and the consumption is just over 3 lb per ton. The charge for each mill is 37½ tons. The mills are driven at 19.2 rpm through a single helical gear and pinion directly connected to a 440 hp synchronous motor.

Ball-mill scoops are of the single type with a 72-in. radius. The ball-mill bay is shown in Fig 2.

Small size steel grinding balls for use in the regrind ball mill are made by the concentrator department. Short lengths of discarded drill steel are heated in a coke furnace. When bright red they are cut in 2-in. lengths in a pair of rotary shears and immediately picked up with tongs and fed to special dies fitted on an I.R. No. 50 drill sharpening machine. One squeeze with this machine converts the red hot, 2-in. length of drill steel to a ball of between 1¼ and 1½ in. in diameter. The ball is then removed from the die and dropped into a tank of water. The whole operation is done with one heating. Old jack bits are treated in the same way.

Life of ball mill liners is given in Table 1.

Classifier overflow gravitates to two pumping stations each of three 8-in. Wilfley pumps.

Because of the various types and capacities of the flotation machines in use a distributor of very flexible design was installed and details are given in Fig 3.

FLOTATION

The flotation plant consists of approximately equal capacities of airlift-type cells and No. 36 Agitair mechanical flotation machines. The latter have proved themselves metallurgically better than the airlifts and also show a saving in power. Mechanical type machines will eventually replace all airlift cells.

The flowsheet is given in Fig 4 and part of the flotation section is shown in Fig 5.

During the past 13 years many different circuits have been tried out on both small experimental scales and on large full scale mill tests.

Present practice is straightforward and simple with roughers, scavengers and two stages of cleaning. The thickening, regrinding and scalping of middlings are conventional practice.

TABLE 4—*Assays of Various Products of Flotation Pulp*

Product	Tons per 24 Hr	Cu Assay, Pct	Cu Distribution, Pct
Heads	8,000	3.40	100.0
Rougher feed	9,487	3.68	128.6
Rougher concentrates	1,247	24.00	109.1
Rougher tails	8,250	10.64	19.5
Scavenger concentrates	736	3.72	9.9
Final tails	7,514	0.34	9.6
Cleaner feed	1,359	23.40	116.9
Cleaner concentrates	531	47.96	93.8
Cleaner tails	828	7.73	23.5
Final concentrates	486	50.87	90.4
Recleaner tails	45	19.11	3.3

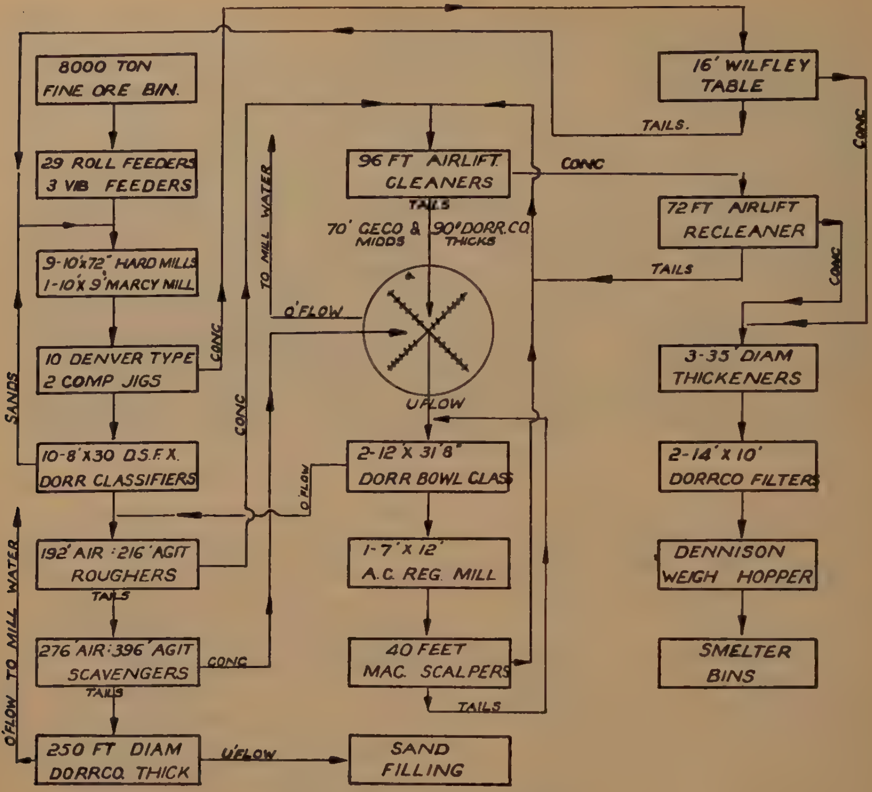

FIG 4—FLOTATION FLOWSHEET.

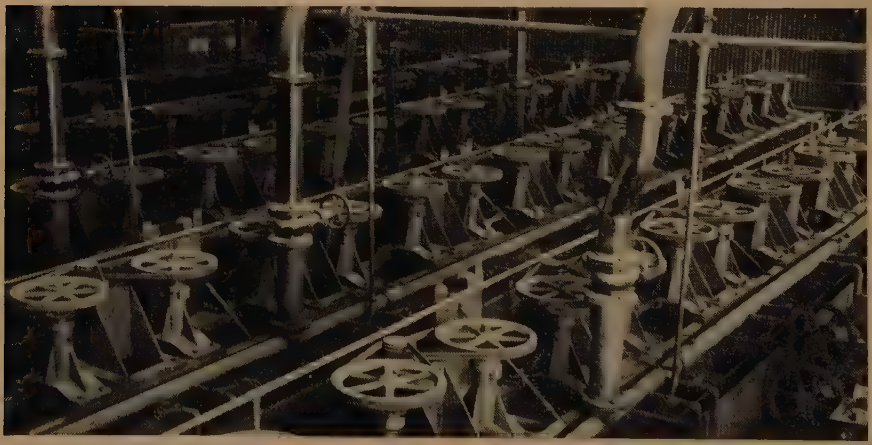

FIG 5—PART OF FLOTATION SECTION.

The density of flotation pulp is in the region of 18 pct solids by weight. Table 4 gives the assays of various products in a typical day's run.

Reagent combination and consumption is as follows:

	Lb per Ton
Sodium xanthate	0.09
Pine oil	0.06
Natal coal tar	0.07
Dextrin	0.04
Sodium sulphide	0.08
Lime	1.73

The use of dextrin as a carbonaceous depressant has proved of great value, particularly when treating predominantly graywacke ore, with its graphitic carbon. The dextrin serves to drop the carbon with its associated slime from the concentrate, thereby increasing the grade to the desired figure.

During the years xanthate consumption has been decreased from 0.75 to the present 0.09 lb per ton.

As mentioned earlier, change of ore in the mill feed has been a disturbing factor in our operation. Xanthate-consuming properties of various types of ore vary widely, clean siliceous quartzite requiring 0.03 lb per ton, and altered inter-ore body schist consuming as much as 3 lb per ton. At times, consumption in the mill had varied by 30 pct within 24 hr, and during a month has ranged between 0.10 and 0.35 lb per ton.

To deal with this variable condition, a routine test was developed whereby the available collector in the tails was determined by actually testing its collecting power on a sample of fresh feed. This test, compared to a standard floated in fresh water, gives us a numerically expressed xanthate content. By keeping a small excess in the tails we are assured of enough in the circuit without undue waste.

Lime concentration has been proved of vital importance to both copper recovery and xanthate consumption. The critical amount for best operation is about 0.07 lb free $Ca(OH)_2$ per ton of water in the rougher tails. This is controlled by titration of a settled sample with standard sulphuric acid and phenolphthalein after carbonates have been precipitated by an excess of barium chloride.

Sodium sulphide is fed in proportion to the tonnage of middlings being circulated. The scalping within the regrind circuit is another vital factor and necessary for good recoveries.

Quick method insol assays and colorimetric copper assays are carried out as part of the routine work of the flotation operators on each shift as a guide to operations.

TABLE 5—*Typical Analysis of Mill Products*

PER CENT

	Cu	CuO	Insol	SiO_2	Al_2O_3	S	Fe	CaO	MgO	Bi
Heads	3.37	0.22	83.02	73.40	8.58	1.22	1.92	3.15	2.27	
Tails	0.37	0.17	88.41	78.43	9.33	0.07	0.87	3.67	2.30	
Concentrates	52.19	0.80	15.20	12.27	2.18	20.33	9.66	0.53	0.52	0.034

TABLE 6—*Recoveries from July* 1945 *to June* 1946

PER CENT

	Cu	CuO	Cu Recovered	CuO Recovered	CuS Recovered	Ratio
Heads	3.414	0.19	100.00	100.00	100.00	
Tails	0.337	0.15	9.29	76.21	5.46	
Concentrates	53.124	0.76	90.71	23.79	94.54	17.15

All pulp is handled by Wilfley pumps.

Air for the flotation machines is supplied by Reavell Turbo blowers operating at a pressure of between 1.3 and 1.5 psi.

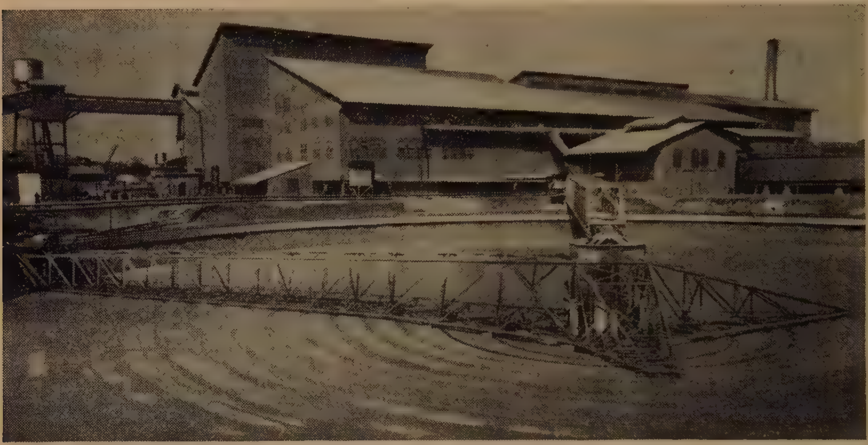

FIG 6—250-FT TAILINGS THICKENER DRAINED FOR REPAIRS.

A typical analysis of the mill products is given in Table 5 and recoveries from July 1945 to June 1946 are shown in Table 6.

Sampling of heads, tails and concentrate is automatic and cuts are taken at between 20 and 30-min intervals.

Mill tailings are thickened in a 250-ft Dorr thickener (Fig 6) to about 55 pct solids and pumped to either cones for sand-filling underground stopes or to one of the tailings disposal sites.

WATER SUPPLY

The clear water overflow from the tailings thickener together with the overflow from two middlings thickeners and three concentrate thickeners gravitate to a central pumping station. Enough fresh water is added here by means of a float valve to maintain a balance and all water is pumped to a steady head tank for the mill supply.

Fresh water is pumped separately to the cleaner circuit for dilution in the final stages of cleaning. This has been found beneficial in helping to drop a little extra insol.

All water pumps are single-stage Mather and Platt pumps.

Lime scale in the water pipes had been a real problem in the past but recently we have been using a threshold water treatment employing Calgon and the results, which are not yet complete, look very promising.

FILTERING

After the flotation concentrates are thickened, they are filtered in two 12-ft diameter by 14-ft long B.M.B.C. Dorrco filters which produce a cake containing about 10 pct moisture at a rate of 20 dry tons per hour each.

Before delivery to the smelter the concentrate is accurately weighed in a Blake Dennison batch weighhopper. This machine is carefully tested monthly in the presence of the Smelter and Mill Superintendents and the Mechanical Engineer and any error found in the test is applied to the tonnage for the month by both mill and smelter. This arrangement works admirably as it leaves no room for argument between mill and smelter and the weighhopper is a constant performer that seldom shows an error of over 0.06 pct.

TAILINGS DISPOSAL

Tailings that are not used for sand filling underground are pumped and gravitated to a disposal site in a flat valley through which runs the Mufulira stream. A clear water overflow is maintained from this dam except when heavy rains wash down a lot of mud and fines that have collected during the dry part of the year (Fig 7). Tailings are also used to advantage to fill in swampy depressions as an anti-malaria measure.

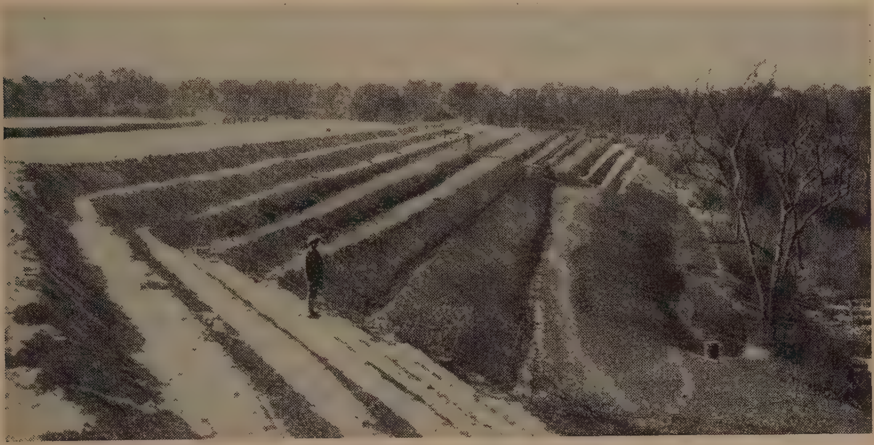

FIG 7—NO. 7 TAILINGS DAM.

LABOR

The crushing plant is operated for 3 shifts of 8 hours each by one shift boss, 2 European operators and 33 Africans. The ball mills and flotation section have one shift boss, 3 European operators and 18 Africans per shift and the filter plant is operated for only two 8-hour shifts with one European and 4 Africans per shift.

Maintenance and repairs are taken care of by a gang of 17 European fitters and 48 Africans under a foreman fitter. The cost of operations per ton of ore milled are shown in Table 7.

TABLE 7—*Costs per Ton of Ore Milled*

SECTION	SHILLING
Crushing, conveying and screening......	0.58
Grinding and classifying................	1.43
Gravity concentration..................	0.01
Flotation..............................	0.66
Concentrate handling and filtering.......	0.05
Tails disposal.........................	0.07
Sampling and assaying..................	0.06
Mill water supply......................	0.03
Concentrator superintendent dept. and European war allowance..............	0.13
Native cost of living allowance..........	0.06
Total................................	3.08

POWER

Alternating current is supplied by the Company's power plant and is distributed to the mill as a 50-cycle three-phase supply in both 3300 and 500 volts.

Consumption is given in Table 8.

TABLE 8—*Consumption of Electricity*

OPERATION	KW-HR PER TON
Crushing, conveying and screening......	2.87
Hardinge ball mills....................	9.54
Classifiers............................	0.39
Ball mill pumps.........................	0.42
Blowers...............................	2.60
Flotation pumps........................	1.30
Flotation machines.....................	0.52
Filtering.............................	0.23
Sampling and assaying..................	0.03
Tailings disposal......................	0.33
Mill make-up water....................	0.76
Lighting..............................	0.05
Total................................	19.04

ACKNOWLEDGMENTS

Thanks are due to the Management who have always encouraged any ideas of the Concentrator organization that are of a progressive nature.

DISCUSSION

(*D. W. Scott and J. F. Myers presiding*)

CO-CHAIRMAN MYERS—The flowsheet and picture show that a Marcy open-grate ball mill was tested out against the main body of Hardinge high level discharge mills. There is little or no data in the published literature upon this subject and I think it would be very interesting if the Mufulira data could be presented. I understand from reliable sources that the test showed the power consumption to be a stand off, but that the ball consumption was some 30 pct less in the high level discharge mill.

The motor on those trunnion discharge mills and the Marcy grate mill were the same size, were they not?

R. B. ADAIR (author's reply)—They were, 400 hp.

W. L. MAXSON[*]—This is an open field for controversy but the findings which the authors report, I think, are pretty well in agreement with the test work that we conducted under somewhat similar conditions.

I think even those who are very strong advocates of the grate type of mill will concede that while more tonnage will go through the mill—and by that I mean a given size mill—there is the possibility of greater capacity, but the power consumption increases. In general, the liner wear and ball wear go up, and I have always thought it could be reduced pretty well to a matter of capital investment.

If we had Walter Garms here this morning I think he would go along with us and tell us if we were to use just an overflow mill and run it slow enough we would get the most for our dollar.

There are places, no doubt, where the grate type mill has definite advantages, and there are places where the overflow has definite advantages. I am inclined to think that you have to consider both types, but I do not think one can adopt either of them to the complete exclusion of the other.

The test data in the paper are not as complete as I would like, and I think it would be of service to the industry if at some time those tests might be presented in their entirety as comparative grinding tests, applying to a particular set of conditions. It would just be another set of notes for the record which might help someone else who is faced with a similar problem.

I do not know whether Mr. Adair is familiar with the preliminary test work conducted on the N'Kana milling operation. Were you familiar with that?

R. B. ADAIR—I have been to an extent, yes.

W. L. MAXSON—You recall they used a mill which was about 6 by 8 ft equipped with means to vary the speed with grates and without grates. The test data led to the installation of overflow type mills. Is that not true?

R. B. ADAIR—That is true.

W. L. MAXSON—That is a rather comprehensive test and those data are available. However, you are always open to question as to whether you would have secured the same results if you had tested two mills side by side on a split feed.

G. GUTZEIT[*]—For the determination of excess xanthate, I have used a method which gave satisfactory results. It consisted in transforming the xanthate present into the molybdenum complex ($MoO_3[SC(SH)(OC_2H_5)]_2$), which has a characteristic deep blue-violet coloration. For that purpose, a measured volume of the solution is treated with acidified ammonium molybdate (2N-HCl). The violet complex is then shaken out in a separatory funnel with a standard volume of an organic solvent, such as benzene or chloroform. Finally, the extract is examined in a colorimeter (electrophotometer). A green light-filter is used. The apparatus has been, of course, standardized previously with known amounts of potassium xanthate.

The sensitivity is fairly high, the concentration limit being 1:1,250,000.

R. B. ADAIR—The method we used is the standard method as suggested by the company that produces the xanthate, Great Western Division, Dow Chemical Co.

R. J. MORTON[†]—With regard to what Dr. Gutzeit had to say about the determination of xanthate in solutions, I think that is quite possible in dilute solutions having a relatively low soluble salt concentration. However, in pulps where the solutions contain as high as 2000 ppm of soluble salts, we found that all methods—including I believe the particular method he has mentioned—were useless. The determination of xanthate ion in solutions from flotation pulps of heavy sulphide ores is made very difficult when the total concentration of soluble salts is of an order 50 to 100 times that of the xanthate ion.

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[*] Oliver Iron Mining Co., Duluth, Minn.

[*] Director, Research and Testing Laboratory, General American Transportation Corp., East Chicago, Ind.

[†] Oliver Iron Mining Co., Duluth, Minn.

F. X. Tartaron*—The pH versus lime comparison goes back about 20 years to my knowledge. In every case that I know of, the decision was that titration was better. The explanation is that when you place lime in water, it is a weak base and only a small percentage will ionize. This ionized portion is measured by pH. If the ore characteristics change and lime is consumed, the ionized portion is removed but is immediately replaced by ionization of the un-ionized portion. Thus the pH can remain constant and yet lime be consumed. Was your titration on tailing water?

R. B. Adair—Yes.

F. X. Tartaron—If you determine the pH you would not get a figure on the actual lime consumption; but if you titrate, comparing the lime actually added to the lime present in the water, you can compute the lime consumption. We made quite a number of such determinations on flotation of Minnesota iron ores and could follow the lime consumption through the various steps of the process. I do not know of any case, perhaps someone else does, where pH was usable as a determinant of the quantity of lime to be employed.

Did you find your titrations useful, the same as you found the determinations of xanthate useful in regulating your flotation?

R. B. Adair—Very much so, yes. They were our chief guide to operation. The method that I described in some detail was on the determination of what we called the collector power, because sometimes we were using other collectors in addition to xanthate. In a case like that your xanthate method would fail.

One reason this particular method was valuable there was that it was fast, it could be done by unskilled labor in a matter of 30 to 40 min. We could determine exactly the condition that was in our cells insofar as their needs, and at other periods we would go back and check the collector condition right through the cells from one end to the other.

This was the factor that made us decide on stage addition of xanthate.

F. X. Tartaron—How did you differentiate between xanthate and lime when you used the tailing water?

R. B. Adair—The methods for xanthate and lime were quite independent of each other. The lime determination was quite unaffected by the concentration of collector. To carry out the quite arbitrary test on collector concentration, it was necessary to first check to see that the lime condition was normal. If necessary, this was first corrected, then a sample taken for the xanthate test.

A. F. Taggart*—Did you get any lead at all on the mechanism by which you lost xanthate with the slime ores?

R. B. Adair—I am afraid, Professor Taggart, I can suggest nothing in that line. We gave it a great deal of thought and discussion but we came to no conclusions worth recording, I would say.

A. F. Taggart—I expected so, but some of the hunches you had and some of the leads you might have had would be, I think, valuable for the record. You are not going to be held down to them at all, but the other fellow's hunches are often valuable.

R. J. Morton—I do not wish to start an argument as to the mechanism of adsorption, but it is possible that the adsorption of xanthate in this particular instance parallels that of gold adsorption by certain talcose slimes. This is particularly noticeable in practice when using counter current decantation for solution recovery in the cyanidation process.

E. H. Bronson†—My comment on this paper is that although it was well prepared, many details had been omitted which would have made it of much more value. For example, in the description of the crushing plant we are unable to find information on the number of hours per day the plant operates or the hourly tonnage capacity. The only clue is the statement that the McCully crusher delivers a 5 in. product at a rate of over 500 tph.

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* Jones and Laughlin Steel Corp., Ishpeming, Mich.

* Department of Mineral Dressing, School of Mines, Columbia University, New York, N.Y.

† Metallurgical and Mechanical Consulting Engineer, Toronto, Ont., Canada.

R. B. Adair—The crushing plant normally operates two shifts, leaving the day shift available for maintenance and clean-up.

The hourly tonnage rate is considered as 500, but wet ore can reduce this materially. The latest figures I have available are for August 1947, in which the rates were as follows:

	Tons per Hour	Number Installed
McCully Crusher..............	450	(2)
Symons Standard.............	125	(4)
Symons Shorthead............	85	(5)

Since the daily tonnage varied from 5500 to 9000, the actual operating time also varied widely.

E. H. Bronson—Table 2 is labelled as typical screen analysis of crushing plant. Is this the product from the final screens or is it the feed to the fine ore bins?

R. B. Adair—Table 2 is the screen analysis of the crushing plant product, that is, the feed to the ball mills.

E. H. Bronson—How were the Denver jigs modified to treat the native copper? What was the method of collecting the jig concentrates and transporting to the table? Would a cleaner jig have worked more satisfactorily?

R. B. Adair—The jigs were modified in several ways, the first being the installation of a third central bearing on the main shaft. Prior to this the whip in shaft was sufficient to wear out bearings in a very short time. It was found that irregular shaped pieces of native copper would pass through the trash screen and blind the lower wedge-bar screen. Thus it was necessary to use a relatively fine trash screen and coarse lower screen. Even with this frequent changes in jig bedding were necessary to keep the jig working.

Jig concentrate was discharged continually from a spigot (5⁄16 to 3⁄8 in.) carried in launders to a Wilfley pump, which discharged into a dewatering cone. The cone spigot fed the table. A cleaner jig was used at first in place of the table, but it was found difficult to control with such a dilute feed. The table gave a visual control of the jigs, in that a dirty hutch product was at once evident on the table, and could be shut down for bed renewal.

E. H. Bronson—Are cast iron or steel balls used? The consumption of 3 lb per ton seems high for a comparatively soft ore at a coarse grind.

R. B. Adair—Most of the balls used were cast iron, though some forged balls were tried. The cast balls came from both Southern Rhodesia and South Africa in addition to those made by the smelter department from scrap steel. Even with balls of Brinell hardness up to 600, the consumption was very high. Mufulira ore is not soft. It is tough and very abrasive. You may be thinking of the ore at Roan Antelope which is relatively soft. Our grind of about 60 pct minus 200 mesh could scarcely be called a coarse grind.

Description of Concentrating Operations, Roan Antelope Copper Mines Limited, Northern Rhodesia

By M. R. Goldick*

(New York Meeting, February 1948)

The Roan Antelope concentrator was originally designed with a nominal milling capacity of 6000 tons of copper ore per day but this was subsequently considerably exceeded. In broad outline the plant consisted of two No. 30 McCully gyratory crushers for coarse crushing, five 5½-ft Standard Symons crushers in open circuit for secondary crushing, ten No. 98 Marcy ball mills arranged in five sections, each two-mill section with primary and secondary Dorr classifiers for two-stage grinding. One stage of flotation roughing was followed by two stages of cleaning, all in airlift type pneumatic flotation machines, with the final concentrates filtered on Oliver drum filters. Tailing was thickened in a 250-ft Dorr traction type thickener before pumping to the tailing dam.

Concentrating operations commenced in June 1931, on the completion of the first two ball-milling sections and output was increased during the next few months as the remaining sections were finished. From the start, very few technical difficulties were experienced but a number of evolutionary changes have been made as a result of operating experience and the availability of improved equipment. These have resulted in plant and ore-flow alterations although the general scheme of ore treatment remains substantially as before. The main changes to date were the installation of a new crushing plant at Storke shaft, some 7000 ft west of the original Beatty shaft; the completion of a two-stage secondary crushing plant; an increase of 20 pct in ball-milling capacity; a change-over from pneumatic to a mechanical type of flotation machine for flotation roughing; the reduction of flotation cleaning from two stages to one stage.

Ore Treated

Typically the ore treated consists of metamorphosed shales and sandstones with shales as the principal mineral-bearing rock carrying chalcocite, bornite and chalcopyrite as the main copper minerals. Chalcocite predominated in the original eastern end of the ore body but there is a pronounced trend to an increased proportion of the copper-iron sulphides as mining proceeds westward. Although the ore can be described as soft, this advantage is offset by the extremely fine dissemination of the copper minerals which calls for a high degree of comminution of the ore (90 pct minus-200 mesh Tyler standard) to give a reasonably satisfactory separation of mineral from gangue. Latterly, there has been a tendency for the ore to become harder as shown by a reduction in the ball-milling rate from a previous figure of over 900 tons per ball-mill day to the present output of some 800 tons. Chalcopyrite is not so finely disseminated and has also proved more amenable to flotation than chalcocite. Nonsulphide minerals are mainly the so-called oxides such as malachite, cuprite and melaconite with some chrysocolla and

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* Concentrator Superintendent, Roan Antelope Copper Mines Limited, Luanshya, Northern Rhodesia.

occasional noticeable amounts of native copper.

A typical present day mill feed assay is: total Cu, 2.60 pct; oxide Cu, 0.12 pct; SiO_2, 55.5 pct; Al_2O_3, 17.2 pct; Fe, 3.9 pct; S, 0.9 pct; CaO, 1.8 pct; MgO, 4.6 pct; undetermined, 13.5 pct. This type of ore calls for no unusual method of treatment and the process in use closely follows accepted flotation concentration practice for sulphide copper ores.

Plant Layout

The main plant arrangement is shown in the photograph (Fig 1) and in the schematic layout (Fig 2) which indicate the compact and convenient disposition of the secondary crushing, milling and flotation sections. The coarse crushing plant is situated at Storke shaft and the broken ore is transported some 7000 ft to the secondary crushing plant by a five-section, 42 in. wide, belt conveyor system.

This is not a gravity flow layout in the generally accepted sense but in this plant, with large quantities of various circulating loads, the benefits to be gained by gravitational flow of all of the main products would be largely offset by the added structural complications of a sloping plant site.

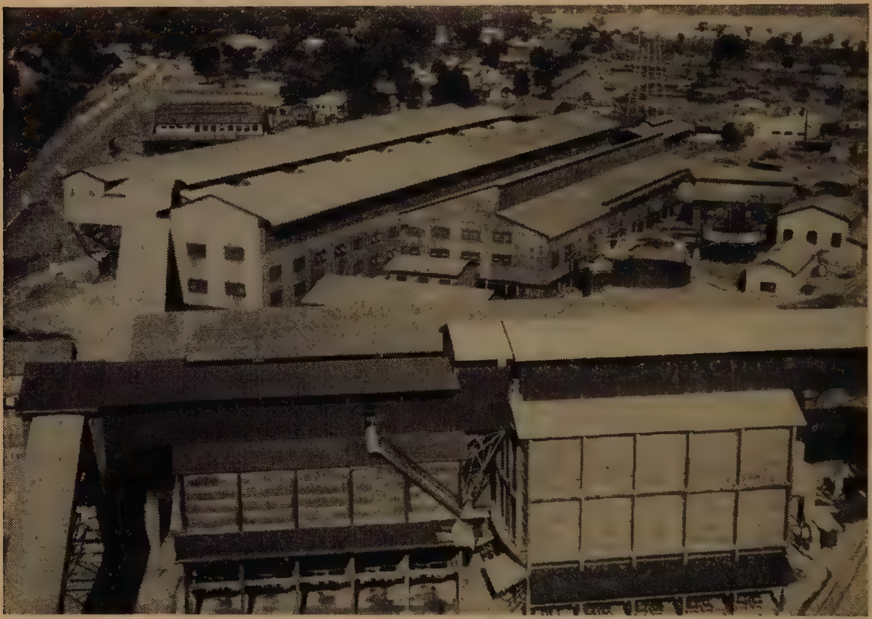


Fig 1—Main plant arrangement of concentrator.

In order to take advantage of cheap transportation of concentrate by pumping, the filter plant is close to the smelter storage bins for flux, charge and reverts. The tailing thickener was so placed as to obtain a gravity flow from the mill and still be adjacent to the dam, thereby making use of only one pumping operation to complete the disposal of the tailing. These two sections are not shown on the plan but are situated 1700 ft and 2500 ft, respectively, in a southeasterly direction from the main concentrator building.

The coarse crushing plant, the secondary crushing plant, the main concentrator building, the various repair workshops and the mine store all have connection to the

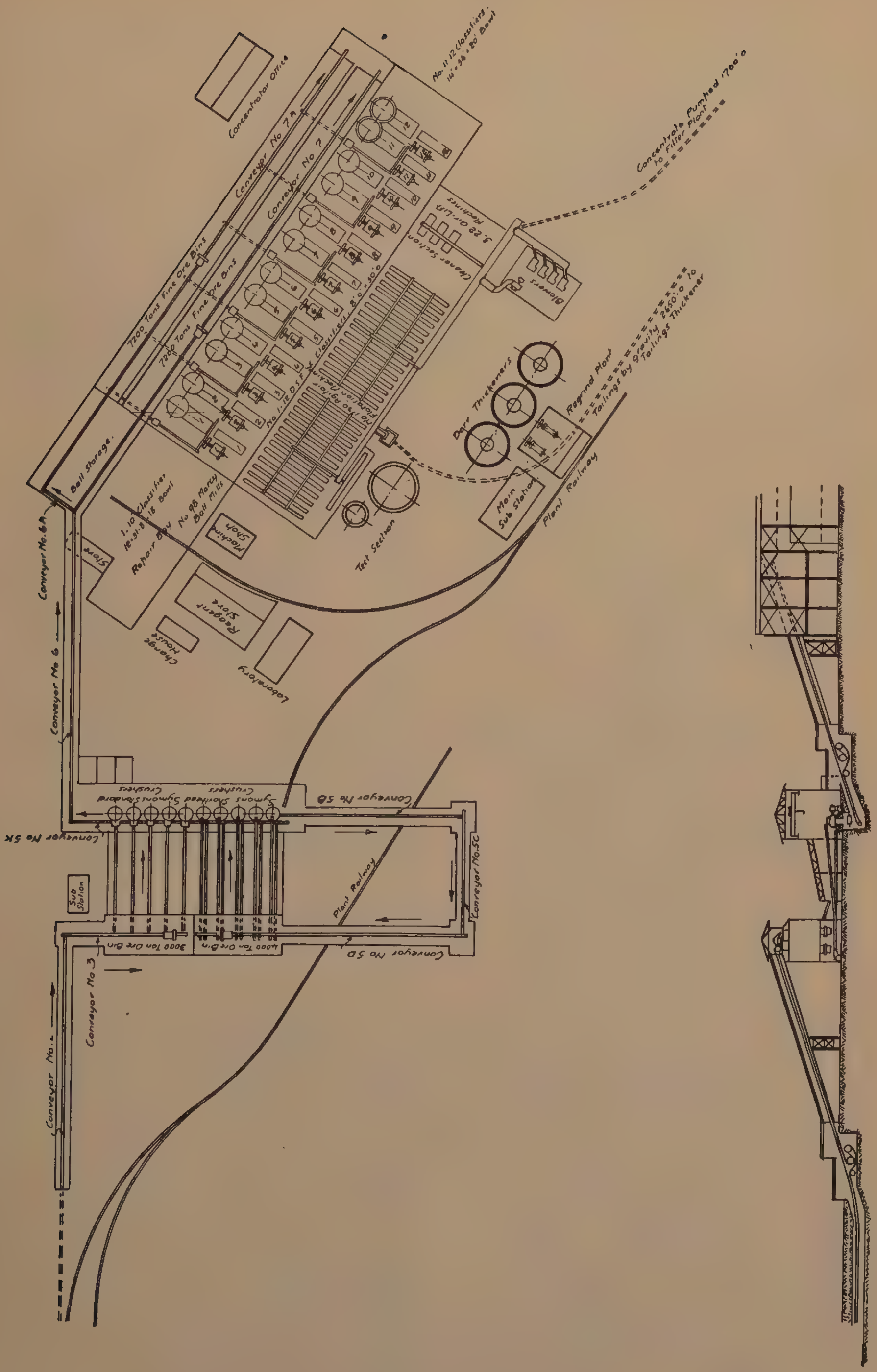


FIG 2—SCHEMATIC LAYOUT OF CONCENTRATOR.

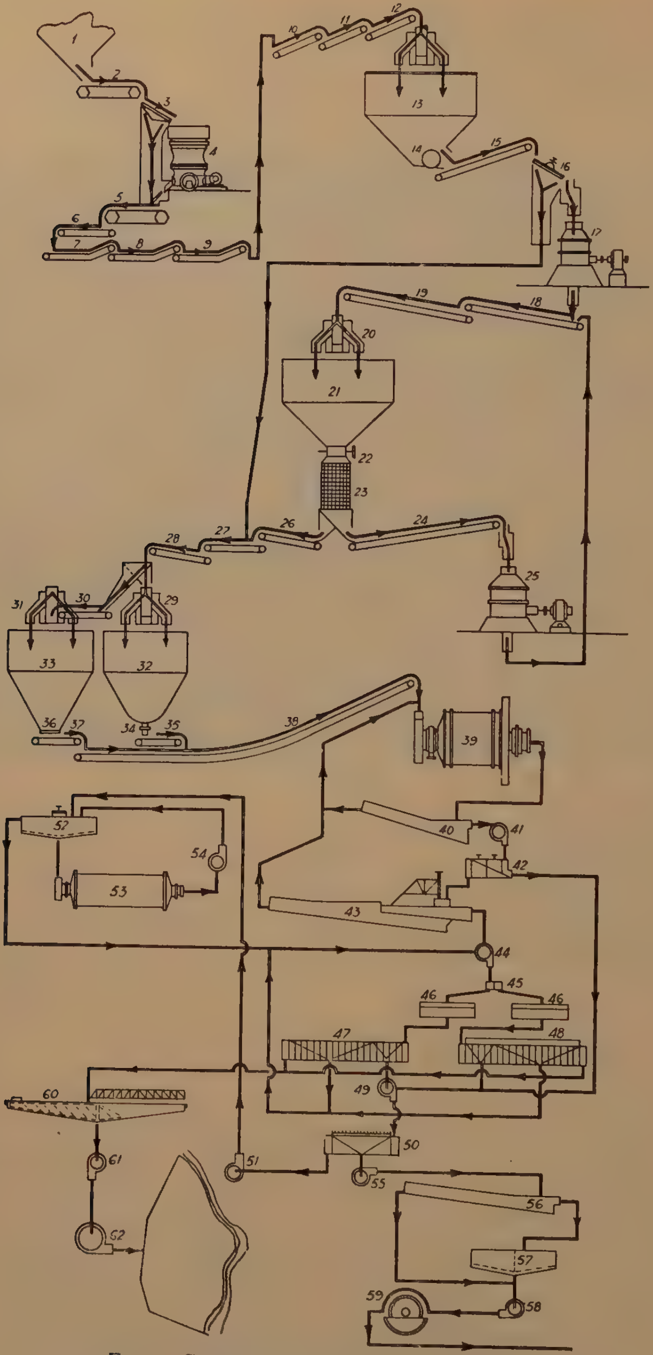


FIG 3—GENERAL FLOWSHEET OF CONCENTRATOR.
(See opposite page for explanation)

1. Storke headframe bins. Capacity 2000 ton.
2. Two 72-ft pan feeders. Variable speed dc motor drive.
3. Two 6 × 10 ft grizzlies. 3½ to 4 in. openings. Hinged at bottom, sprung at top.
4. Two 30-in. McCully gyratory crushers. 180 hp motor drive through 14 1 × 1¼ in. *V* belts.
5. Two 48-in. pan feeders. Variable speed dc motor drive.
6. 42-in. belt conveyor. 400 ft per min. Length, 377 ft: rise, 23 ft 6 in. 50 hp motor: No. W1.
7. 42-in. belt conveyor. 400 ft per min. Length, 2088 ft: fall, 17 ft 4 in.: 125 hp motor: No. W2.
8. 42-in. belt conveyor. 400 ft per min. Length, 2096 ft: fall, 37 ft 6 in.: 125 hp motor: No. W3.
9. 42-in. belt conveyor. 400 ft per min. Length, 1548 ft: rise, 7 ft 6 in.: 125 hp motor: No. W4.
10. 42-in. belt conveyor. 400 ft per min. Length, 1285 ft: rise, 18 ft 5 in.: 125 hp motor: No. W5.
11. 42-in. belt conveyor. 400 ft per min. 150 hp motor: No. 2
12. 42-in. belt conveyor. 400 ft per min. Length, 100 ft with automatic tripper: No. 3
13. Standard Symons storage bins. 3000-ton capacity.
14. Five 48-in. drum feeders.
15. Five 28-in. belt conveyors (No. 4A to 4E) incline lift 27 ft. Magnetic head pulley. 15 hp motor.
16. Five 4 × 6 ft Type 60 Hummer screens set at 37° with ½ × 1¼ in. openings.
17. Five 5½-ft Standard Symons cone crushers. Direct coupled to 150 hp motor.
18. Main 48-in. collecting belt conveyor (No. 5B). 150 hp motor.
19. One 48-in. cross conveyor (No. 5C). 150 hp motor.
20. One 48-in. conveyor with automatic tripper (No. 5D). 150 hp motor.
21. Shorthead Symons storage bins. 4000-ton capacity.
22. Five sets of two 30-in. drum feeders—5 hp motor for each set.
23. Five sets of two 60 × 102 in. double deck Gyrex screens. Top cloth 1 × 1 in. Bottom ⅜ × 1¼ in. openings.
24. Five 30-in. inclined feed conveyors (No. 4F to 4K) 15 hp motors. Magnetic head pulleys.
25. Five 5½-ft. Shorthead Symons crushers. *V* belt drive from 200 hp synchronous motors.
26. Five 24-in. conveyors for Gyrex screen undersize (No. 5E to 5J) 10 hp motors.
27. One 42-in. collecting belt (No. 5K). 25 hp motor.
28. One 42-in. conveyor (No. 6). Incline lift of 75 ft. 125 hp motor.
29. One 42-in. conveyor (No. 7) with automatic tripper. 75 hp motor.
30. One 42-in. cross conveyor from splitter (No. 6A). 20 hp motor.
31. One 42-in. conveyor with automatic tripper (No. 7A). 70 hp motor.
32. No. 1 fine ore bin. Capacity 7200 tons in 6 sections of 1200 tons.
33. No. 2 fine ore bin. Capacity 7200 tons in 6 sections of 1200 tons.
34. Thirty-six 18-in. roll feeders. Six feeders per bin section. 5 hp motor per set of six feeders.
35. Six 20-in. belt conveyors (No. 8), 46-ft long, collecting from roll feeders. 5 hp motor each.
36. Thirty-six 3-ft wide pan feeders. 6 feeders per bin section. 5 hp geared motor per feeder.
37. Six 24-in. belt conveyors (No. 8A) collecting from pan feeders. 5 hp motor each.
38. Six 20-in. belt conveyors (No. 9) Merrick Weightometer on each. 10 hp motor each.
39. Twelve 9 × 8 ft Marcy ball mills. 20.3 rpm. Direct-coupled to 400 hp motors.
40. Twelve 8 × 30 ft Dorr FX classifiers. 20 strokes per min. 25 hp motors.
41. Twenty-four 6-in. Wilfley pumps. Two per mill. One in use with other as stand-by.
42. Twelve 2-cell scalping flotation machines. 24-in. impellers. Concentrates to cleaner flotation machines.
43. Ten Dorr quadruplex bowl classifiers. 12 × 32 ft with 18-ft bowl. 15 hp motors. Two Dorr FX turret bowl classifiers. 14 × 36 ft with 20-ft bowl. 25 hp motors.
44. Four 8-in. Wilfley pumps. Bowl classifier overflow to flotation sections.
45. One 2-way distributor.
46. Two 15-way distributors supplying mechanical flotation machines.
47. Fifteen 18-cell Agitair flotation machines. 5 hp motor for each two cells. Roughers.
48. Fifteen 18-cell Agitair flotation machines. 5 hp motor for each two cells.
49. Two 6-in. Wilfley pumps. Primary rougher concentrates to cleaner section.
50. Three 22-ft airlift flotation machines as cleaners.
51. Two 6-in. Wilfley pumps. Intermediate flotation products to regrind sections.
52. Two 35-ft diam Dorr thickeners modified for use as classifiers.
53. One 4-ft 6-in. × 12 ft and one 7 × 12 ft ball mill. Grinding intermediate flotation products.
54. One 6-in. Wilfley pump elevating mill discharge to classifiers.
55. Two 3-in. CA Wilfley pumps. Final concentrate to filter plant.
56. One 6 ft × 20 in. Dorr classifier. 12 strokes per min. 5 hp motor. Coarse product to filter.
57. One 50-ft Geco thickener. 2½ hp motor. Thickened product to filters.
58. Two 3-in. CA Wilfley pumps. Thickener underflow to filters.
59. Three 8 × 8 ft Oliver drum type filters. Cake 10 pct moisture to smelter bins.
60. One 250-ft Dorr traction thickener. 2 rph. 10 hp motor.
61. Four 6-in. Wilfley pumps. 125 hp motors.
62. Two 8-in. Wilfley pumps. 125 hp motors. Delivering to main tailings dam.

standard 3 ft-6 in. gauge railway line which enables heavy spares or parts for repair to be transported and off-loaded where required.

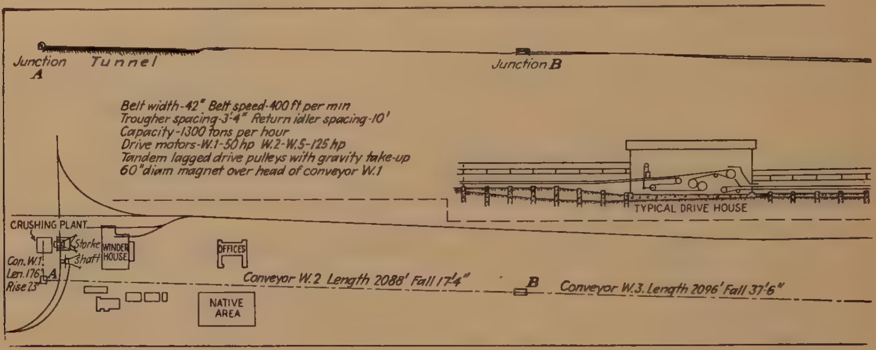

Junction A
Tunnel
Junction B
Belt width-42" Belt speed-400 ft per min
Trougher spacing-3'-4" Return idler spacing-10'
Capacity-1300 tons per hour
Drive motors-W.1-50 hp W.2-W.5-125 hp
Tandem lagged drive pulleys with gravity take-up
60" diam magnet over head of conveyor W.1
TYPICAL DRIVE HOUSE
CRUSHING PLANT
Storke Shaft
WINDER HOUSE
OFFICES
Con. W.1. Len. 176' Rise 23'
A
Conveyor W.2 Length 2088' Fall 17'-4"
B
Conveyor W.3. Length 2096' Fall 37'-6"
NATIVE AREA

FIG 4—SURFACE ORE-

A general flowsheet of the concentrator is shown in Fig 3 and this also gives a large amount of data covering the equipment in use.

COARSE CRUSHING

The coarse crushing plant is situated at the side of Storke shaft headframe where ore skips dump into storage bins of 2000-ton capacity. All ore is passed through 18-in. spaced grizzlies before loading into skips but occasionally large slabs occur and runs of wet fines cause difficulties. All crushing equipment is in duplicate.

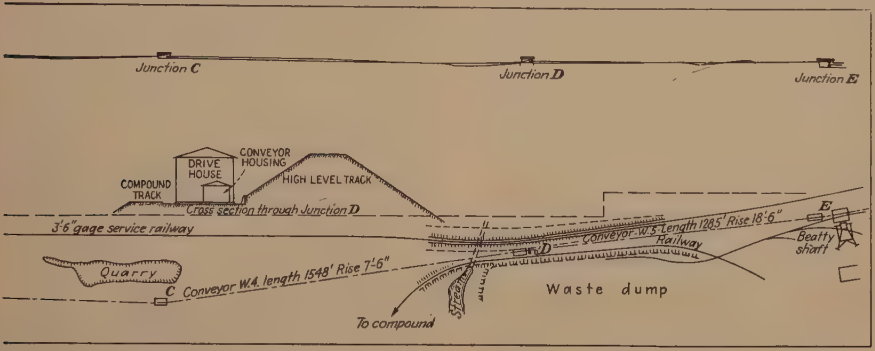
Ore is removed from the bin by 72-in. wide, heavy-duty, variable speed, apron feeders delivering onto spring supported grizzlies. The grizzly oversize passes to No. 30 McCully gyratory crushers set to about 4 in.; the crusher discharge and the grizzly undersize both drop onto 48 in. wide, variable speed, apron feeders which deliver to the first section of the surface ore transport system. The 72-in. apron feeders provide a positive and controlled rate of feed to the crushers while the 48-in. feeders protect the conveyor belt from the severe treatment that it would otherwise receive from large pieces of ore dropping onto it from the crusher discharge and the grizzly undersize. The speed of the smaller feeders is so set that they discharge at about the average crushing rate—a certain amount of space above the feeders serving as a storage for temporary surges and permitting a more regular delivery of ore to the belt conveyor below. The use of apron feeders has been amply justified by the satisfactory service given; maintenance has been confined to the replacement of a few links and supporting rollers.

Grizzly spacing was originally set to vary from 2½-in. wide at the top to 3½-in. wide at the bottom, but this has since been increased by an additional inch to minimize choking when handling wet ore. The use of spring supports for the grizzly bars materially assists in reducing the shock of large blocks of ore dropping onto the bars but has not proved of any particular value in preventing chokes caused by damp fines in the ore. The two McCullys have a capacity of about 500 tons per hour each and are frequently in operation together, giving a total tonnage of upwards of 1200 tons per hour of initial feed to the grizzlies which is close to the rated capacity of the surface ore transport system.

SURFACE ORE TRANSPORT SYSTEM

The surface ore transport system (Fig 4) which carries the crushed ore from the primary crushers to the secondary crushers is not part of the concentrator proper but the association is so close that it is considered that a description of this system can well be included here.

This system consists of five individual tandem belt conveyors covering the 7000 ft from Storke shaft coarse crushing plant to the secondary crushing plant adjacent to the concentrator. There is a fall of about 6 ft in favor of the load with the lowest point some three-fifths of the distance from Storke shaft. Preliminary calculations showed that it would be possible to use four long individual conveyors with the lengths so arranged that the power requirements would be approximately the same for each section. However, because of the location of a service shaft and the hoist house, it was necessary to use a short section of conveyor (Conveyor No. W.1) at right angles to the main conveyor line to avoid these obstacles. This conveyor has a 64-in.-diam ironclad type electromagnet suspended over the bronze head pulley to remove tramp iron and steel from the ore.



TRANSPORT SYSTEM.

Conveyor belt width is 42 in. and the belt speed is 400 ft per minute, giving a maximum capacity of 1300 tons per hour. Troughing idlers spaced at 3-ft 4-in. centers, are of 6-in. diam, three section and fitted with taper roller bearings. Rubber covered troughing idlers of the same general design are used under the loading chutes to minimize the impact of large pieces of ore falling onto the belt. Return idlers are of the plain tubular type, 6-in. diam, also fitted with roller bearings but spaced at 10 ft centers. Mechanical power transmission for each drive is from the motor through heavy duty Falk reduction gears. The short conveyor (No. W.1) has head pulley drive to the belt but each of the other longer conveyors is driven through tandem lagged pulleys situated close to the head pulleys.

All five drive motors operate at 3300 volts, three phase ac; the motor for Conveyor No. W.1 is of 50 hp but the other four motors are interchangeable and each of 125 hp. Starting is by remote control from the secondary ore bins; Conveyor No. W.5, is started up first and when up to speed a centrifugal switch energizes the starter for Conveyor No. W.4 and so on progressively along the system.

Various types of conveyor belt were used with the object of obtaining information as to the relative merits of the various makes and types. Operating results indicate that belts of 6-ply, 42-oz duck 3/16-in. top rubber cover give the most satisfactory service closely followed by belts of 9 to 7-step ply, 32-oz duck and 3/16-in. top rubber cover. It is anticipated that with these belts it will be possible to transport 20 million tons of ore before replacement will be necessary.

Maintenance has been extremely light and the very successful operation of the system is undoubtedly a result of the substantial steel work carrying the conveyor idlers and the use throughout of antifriction bearings. Table 1 shows general operating data.

TABLE 1—*General Operating Data, to December* 1945

System started up, Sept. 14, 1939

Tons of ore transported to date	13,859,740
Total running time, hr	21,148
Average tons of ore transported per month	182,400
Average tons of ore transported per operating hour	655
Power consumed, kw-hr per ton of ore transported	0.62
Tons ore transported per man shift	397
Cost in shillings per ton ore transported	0.03
Average operating labor requirements:	
Africans, per shift	6
Repair labor per day:	
Europeans	1
Africans	3

FINE CRUSHING

Fine crushing is carried out in two stages; the first stage in open circuit and the second stage in closed circuit to produce a final product of minus-⅜ in.

Ore from the surface ore transport conveyors delivers onto a 42-in. wide belt conveyor which elevates the ore 72 ft and, in turn, delivers to a distributing conveyor passing over the secondary (Standard Symons) ore bins of 3000-ton capacity.

Five 5½-ft Standard Symons crushers are used in the first stage of fine crushing. Ore for the crushers is taken from the bins by five roll feeders, each of which is driven from the tail pulley of the 28-in. wide belt conveyor that supplies each crusher. Protection against tramp steel is provided for each crusher by a 42 in. diam iron-clad type electromagnet that is suspended over the bronze conveyor head pulley. A type 60 Hummer screen (area 6 × 4 ft) is set at 37° ahead of each crusher and is normally fitted with screen cloths of ½ × 1¼ in. openings which remove approximately 15 pct of the original feed as undersize. Latterly, because of an increased proportion of moisture in the fines in the ore, it has been found advantageous to increase the screen mesh to 1 in. wide so as to minimize screen blockages and delays.

The five secondary fine crushers are also 5½-ft Symons but of the Shorthead type. All ten crushers are in line and the crusher product is elevated to the top of the 4000-ton Shorthead Symons ore bins by a series of three 48 in. wide conveyor belts, the last of which has an automatic tripper for ore distribution.

The Shorthead Symons ore bin has five equally spaced sets of two discharge openings; each pair of openings has roll feeders driven by 10 hp motors through worm reduction gears. Feed rate is varied by means of arc gates, each feeder discharging onto a No. 60-102 (screen size in inches) Gyrex screen which makes a finished minus ⅜-in. product and an oversize that constitutes the feed to the Shorthead Symons crushers. The minus product from the screens, together with the Hummer screen undersize, is transported by belt conveyors to the fine-ore storage bins supplying the ball mills.

The Standard Symons crushers are direct driven by 150 hp motors through flexible couplings but the original drives for the Shorthead Symons crushers consisted of *V* belts from a small motor pulley to a large crusher pulley. This latter arrangement has been found unsatisfactory because of pulley breakage on the crushers, which is liable to happen if there is a sudden choke; slower speed motors with direct drive are now standard on all crushers.

Because of the flaky nature of the ore, which is conducive to dust production, and the almost complete absence of atmospheric humidity during the dry season, the problem of dust control is of particular significance. So-called atomizing sprays have been found quite useless for preventing the dispersion of dust but heavy wetting of the dry ore following the secondary crushers has been of considerable assistance in preventing dusting at the conveyor transfer points and above the ore bins. The Shorthead Symons crushers which handle dry, clean ore are relatively heavy dust producers but these crushers have been effectively hooded and the hoods connected to an exhaust system. All conveyor transfer points, ore chutes and crusher discharge chutes are connected to exhaust systems that draw through dry cyclone-type collectors followed by wet spray collectors and wet scrubbers before discharging to atmosphere. The accumulated dust from the various collectors plus a relatively large quantity of the finer floor sweepings are washed into a pump sump and pumped across to one of the grinding section classifiers by a 2-in. Wilfley pump. This effectively removes the dust and very satisfactory working conditions are maintained, particularly in the main operating areas.

With a regular supply of ore from the mine, the crushing plants have capacity in excess of 800 tons per hour and it is general practice to run these sections on afternoon and night shifts only, leaving the whole of day shift available for general maintenance. When crushing at the rate of 8000 to 10,000 tons per day the labor force consists of one general crusher foreman, two shift foremen, six operators (one each per crushing shift on coarse crushers, Symons crushers and conveyors respectively), three repairmen, one greaser-craneman and a total of 110 Africans.

Costs have shown a general increase during the last year as a result of higher wage rates and the greater cost of all supplies. A typical cost distribution over the various crushing sections described is as follows: labor, 0.08s; power, 0.05s; maintenance, supplies and the like, 0.16s; total, 0.29 shillings per ton milled.

GRINDING AND CLASSIFICATION

The ball milling section was originally laid out for two-stage grinding using No. 98 Marcy ball mills for both primary and secondary grinding. With this arrangement considerable difficulty was experienced in maintaining a correct balance of loads between the two mills and finally this scheme was abandoned in favor of single-stage grinding which has been found more flexible and generally more efficient.

Two parallel sets of fine ore storage bins are provided holding a total of 14,400 dry tons of ore, about 75 pct of which is live load although this proportion varies appreciably with the moisture content of the ore. This tonnage is sufficient to run the plant through Sunday when no ore is hoisted or crushed.

No. 1 ore bin is of the catenary type but No. 2 is of the braced, hopper-bottom type and each bin is divided into six sections with each pair of sections feeding two ball mills. Ore is discharged from No. 1 fine-ore bin by small roll feeders driven from a countershaft running the length of the bin and although these feeders are satisfactory on dry ore they do not give satisfactory clearance when damp fines are present and considerable difficulty is experienced in keeping the ore on the move. To remedy this, each section of No. 2 fine-ore bin is fitted with six pan feeders, each under a 3 × 6 ft discharge opening and this has been found most effective in making possible a practically complete clearance of the bins.

Ore from the feeders falls onto the collecting belts—one per bin section—which in turn deliver to the mill feed belts. The feed belts pass over Merrick Weightometers, for tonnage record purposes, and at the head end a splitter divides the feed equally between a ball-mill scoop box and a short cross conveyor belt that supplies the next ball mill.

Reference to the flowsheet will show that two-stage classification is used. The mill discharge flows down a launder to a 8-ft wide by 30-ft long Dorr FX type classifier and by using a 72-in. radius scoop on the ball mill it has been possible to avoid the use of any auxiliary elevating scoops for the primary classifier circuit. The classifier sands return to the mill by gravity. A 6-in. Wilfley pump elevates the primary classifier overflow to a scalper flotation machine followed by the secondary bowl classifier. The scalper flotation machine consists of a two-cell, No. 7 size, minerals separation machine with two 24-in. diam impellers driven by one 10 hp motor through *V* belts. Approximately 60 pct of the recoverable copper is removed by the scalpers in the form of a 48 pct copper grade of concentrate which is pumped to the cleaner flotation section. Tailing from the scalper flows directly into the bowl compartment of the Dorr bowl classifiers.

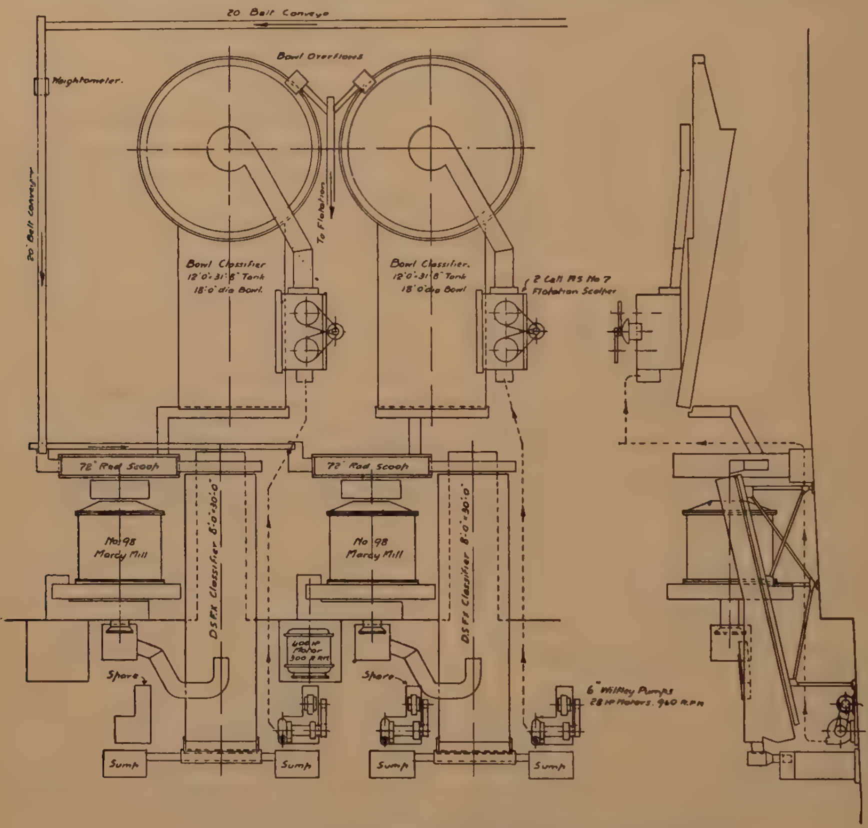


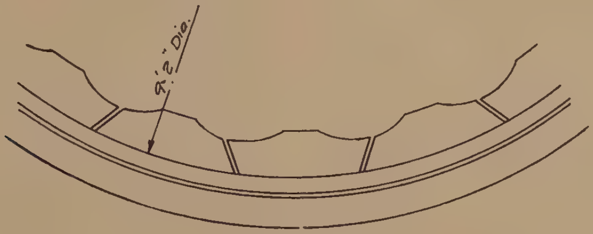
FIG 5—LAYOUT OF GRINDING AND CLASSIFICATION SECTIONS.

The original ten grinding sections had the old style Quadruplex Dorr bowl classifier with a 12-ft wide by 32-ft long raking compartment and an 18-ft diam bowl. These classifiers have continued to give excellent service with the minimum of maintenance but the two newer sections (No. 11 and 12) are equipped with the more modern heavy-duty Dorr FX Turret bowl classifiers with 14-ft wide by 36-ft long raking compartments and 20-ft diam bowls. These can handle heavier circulating loads and give more effective classification.

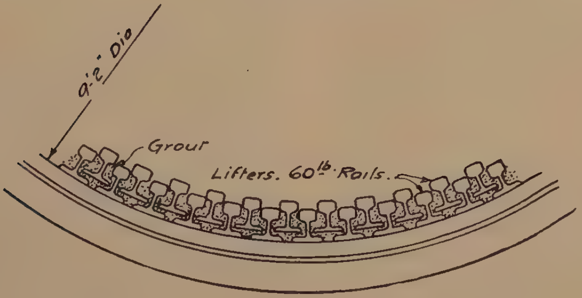
The disposition of the major units of equipment in the grinding and classification sections is shown in Fig 5.

Ball mill discharge is held at between 78 and 84 pct solids as this range covers the most satisfactory pulp density for efficient grinding. The circulating load is about 400 pct of the new feed to the mills with the primary classifier responsible for 150 pct and the secondary classifier for the remaining 250 pct. Final grinding product is maintained at 90 to 92 pct minus-200 mesh (Tyler Standard) and at this fineness the ball mills have a capacity of some 800 tons per ball mill day, varying with hardness of the ore milled.

Steel balls 4-in. diam are used for grinding and these are cast at the smelter from scrap steel melted in an electric furnace. Ball consumption is at rate of 1.4 lb per ton milled and this is less than 10 pct higher than with best quality forged steel balls while the cost per ton milled is less than one third that of the forged balls.



CAST LINERS



RAIL LINERS

FIG 6—LININGS FOR MARCY MILLS.

The ball mills are lined with 60 lb secondhand railway rails (Fig 6). Two interlocking layers of rails are used laid longitudinally against the mill shell and firmly held together in place by ramming a "dry" cement mixture (1 to 2) into the spaces between the rails. One such lining, costing about £50, has a life of 150,000 tons milled compared with 280,000 tons obtained in the past with orthodox cast alloy steel liners costing £1100 per set (pre-war cost—probably at least £1800 now). Feed end liners are cast steel segments, also cast from scrap at the smelter, but discharge grates are still obtained from outside sources. Liner consumption averages 0.20 lb of steel per ton of ore milled.

The whole of the ball mill and repair bay is served by a travelling electric crane with a 90-ton main lift and a 15-ton auxiliary lift. The crane is capable of lifting a fully loaded mill or a complete classifier. A smaller travelling crane serves the whole of the secondary classifiers.

Two spare mill shells are available; one is kept ready lined and loaded with balls for replacement of a mill needing repairs. The second mill is necessary to avoid the possibility of a mill not being available when required, as relining a mill with rails takes upwards of three days although actual replacement of a mill by a relined mill is accomplished in approximately 6 hr. The rail lining is placed in position with the mill shell on end on the floor but for other repairs the shell is placed on a substantial rack with the shell resting on rollers enabling it to be turned as desired and every part made accessible from a convenient working platform. A typical sizing analysis of the mill feed and final product is given in Table 2.

TABLE 2—*Typical Sizing Analysis of Mill Feed and Final Product*

Size	Mill Feed, 3 Pct Moisture		Flotation Feed, 25 Pct Solids	
	Percentage Weight	Percentage Weight Cumulative	Percentage Weight	Percentage Weight Cumulative
on 0.371 in.	13.8	13.8		
3 mesh	24.1	37.9		
6 mesh	18.5	56.4		
20 mesh	14.9	71.3		
100 mesh	6.1	77.4	0.1	0.1
150 mesh	1.2	78.6	1.7	1.8
200 mesh	1.5	80.1	6.8	8.6
−200 mesh	19.9	100.0	91.4	100.0

Labor requirements for operating 12 mills are 2 operators and 12 natives per shift with one shift foreman who is also in charge of the other non-crushing sections of the plant. Cost distribution of grinding and classification per ton milled, on a basis of 8000 to 10,000 tons milled per day, is as follows: labor, 0.08s; power, 0.26d; grinding material, 0.11s; maintenance, supplies and the like, 0.08s; total, 0.53s.

FLOTATION

Because the sulphide minerals are very finely and uniformly disseminated through the ore, the main problem is essentially that of freeing these minerals. Microscopic examination of flotation tailing shows that 85 to 90 pct of the sulphides are in the form of attachments even when grinding to 90 pct minus-200 mesh. The relatively high sulphide-iron ore from the western section of the mine contains an increased proportion of chalcopyrite which is freed at a much coarser degree of comminution than is required for the predominantly chalcocitic ores but, as a considerable proportion of the latter ore will also be milled, it will still be necessary to grind to the usual fineness. The operation of the scalper flotation machines, mentioned when describing the grinding section, has been of considerable assistance in removing the chalcopyrite and bornite as soon as freed at floatable size and thus preventing the overgrinding of these minerals.

The intimate association of mineral and gangue also dismisses the possibility of coarser grinding followed by bulk flotation to produce a low grade concentrate which could be reground for further elimination of unwanted gangue. Present practice is to provide a flotation feed ground to 90 to 92 pct minus-200 mesh and with an ore containing 2.6 pct total copper and 0.12 pct oxide copper. An overall recovery of 91 to 92 pct copper can be made in a concentrate containing 50 to 52 pct copper.

All flotation was originally carried out in airlift type flotation machines with a total length of 220 ft for primary roughing, 660 ft for secondary roughing, 110 ft for cleaning and 80 ft for recleaning. Results were good and maintenance costs were low with the airlift machines but control was not as close as desirable and in 1937 the first mechanically agitated type of flotation machine was introduced and by 1941 all of the rougher flotation section had been changed to this type.

Flotation feed consists of the combined bowl classifier overflows and ranges between 1 and 1.2 pct copper with an original mill feed containing 2.6 pct copper—the reduction in grade of feed being due to concentrate removed by the scalper flotation machines. The flotation feed is elevated by 8 in. Wilfley pumps to a two-way splitter, each section of which supplies a 15-way distributor feeding the rougher flotation machines. All flotation roughing is carried out in a total of 30, 18-cell, No. 36 Agitair flotation machines (cross section of each cell, 36 × 36 in.); each pair of cells being *V* belt driven from a 5 hp vertical electric motor. The first four feed-end cells produce a primary rougher concentrate of 24 to 30 pct copper grade which is pumped to the cleaning section. Secondary rougher concentrate, from the remaining 14 cells in each machine, averages about 2.5 pct copper and is returned to the feed end of the rougher machines. Tailing from the last cell of each rougher machine flows 2500 ft by gravity to the tailing thickener.

The mechanically agitated flotation machines have given improved operating efficiency and better metallurgical results but maintenance costs are appreciably heavier than with the airlift machines.

Flotation cleaning has been reduced to one stage only, using two rebuilt 22-ft long, airlift type flotation machines for this purpose with a third machine available as a stand-by. Feed to the cleaning section consists of primary rougher concentrate plus the concentrate from the scalper flotation machines. The latter can make a finished grade of concentrate but for control purposes it is more convenient to make finished concentrate at one position only. With the inclusion of the scalper concentrate, the feed to the cleaner machines ranges from 35 to 40 pct copper and a finished grade is produced of 50 to 52 pct copper which is pumped 1700 ft to the filter plant. Cleaner tailing averages 10 pct copper and is pumped to the regrind ball mill section. This section consists of a 35-ft diam Dorr thickener for preliminary dewatering; water to the mill water system and thickened underflow to a second 35-ft thickener which has been adapted to serve as a hydroclassifier by reducing the tank height and raising the rake speed. The underflow from the classifier passes to one 4½ × 12-ft ball mill and one 7 × 12-ft ball mill operating in parallel. Mill discharge is pumped back to the classifier, the overflow from which constitutes the final regrind section product and is returned to the main flotation feed.

Although the cleaner tailing is reground to 98 pct minus-325 mesh there is still an appreciable quantity of mineral attachment present and some difficulty is experienced in making a satisfactory grade of concentrate with reasonable recovery. Separate treatment of the reground cleaner tailing was tried for some time but the best overall results have been obtained by returning this material to the main flotation feed.

A typical analysis of final concentrate is: total Cu, 51.6 pct; oxide Cu, 0.6 pct; SiO_2, 14.9 pct; Al_2O_3, 4.1 pct; Fe, 6.9 pct; S, 18.0 pct; CaO, 0.4 pct; MgO, 0.6 pct; undetermined, 3.5 pct.

Air for frothing is supplied at 1.4 psi by two positive displacement blowers each driven by a 125 hp motor and delivering a total of 25,000 cu ft of free air per minute. These are to be replaced shortly by a turboblower with a capacity of 30,000 cu ft of free air per minute.

Flotation reagents currently consist of potassium (or sodium) ethyl xanthate, 0.035 lb per ton; pine oil, 0.55 lb per ton; sodium sulphide, 0.130 lb per ton of ore milled. The first two are added to the ball-mill circuit and the last is added ahead of the flotation feed pumps. Reagents are stored and mixed in a separate building; the mixed reagents and the pine oil are elevated by air pressure through separate pipes to the appropriate feeders which are of the disc and bucket type.

Both the mill-circuit water and the make-up water have a slight natural alkalinity (pH 7.8 to 8.1) and no lime or other conditioning agent has been found either necessary or desirable.

Wet samples are taken automatically by electrically operated and controlled samplers that cut at 20 min intervals and all samples are dried to constant weight in a thermostat-controlled, electric drying oven.

Flotation labor per shift consists of one operator for roughing, one operator for flotation cleaning and sampling and a total of 5 Africans. Costs per ton milled are divided as follows: labor, 0.04s; reagents, 0.13s; royalties, 0.17s; power, 0.08s; supplies and the like, 0.03s; total, 0.45 shillings.

Concentrate Handling

Cleaner concentrate at 50 to 55 pct solids is pumped 1700 ft to the filter plant using 3 in. Model CA Wilfley pumps driven by 40 hp motors, one pump usually being sufficient. These pumps are a special high-pressure, heavy-duty type with feed inlet at shaft height and have given very good service.

Originally, the concentrate was thickened to about 70 pct solids in a 50-ft diam "Geco" traction-type thickener and then pumped to the filters. With the production of concentrate by the scalper flotation machines the final concentrate increased in coarseness from 98 pct minus-200 mesh to the present figure of around 82 pct minus-200 mesh. The coarser product settled rapidly and this, together with an increase in concentrate production, tended to overload the thickener. To rectify this, a 6-ft wide Dorr duplex classifier was rebuilt to the low slope of 1¼ in. per foot and set ahead of the Geco thickener. The classifier operates at 11 strokes per minute and, with the concentrate first fed to the classifier, approximately 60 pct is removed in a relatively coarse product. The classifier overflow gravitates to the thickener and the thickened product joins the classifier sands and both are pumped to the 8-ft diam by 8-ft long Oliver drum filters. Three filters are installed but, with a new cloth, one filter is capable of drying over 300 tons of concentrate per day from an initial feed of 70 pct solids to produce a filter cake of 9 to 10 pct moisture. A series of three 20 in. wide conveyor belts deliver the filter cake to three stock bins of 125-ton capacity each, from which the concentrate is removed by belt feeders into the smelter concentrate handling system as required.

The thickener overflow and the filtrate from the filters pass to brick settling vats which recover a certain amount of very fine suspended concentrate and then the clear water flows to the main mill water supply tanks.

Because of the relatively large amount of free alumina in the concentrate, the filter cloths blind rather rapidly and the usual life of a cloth is one month with one removal and washing during that period.

In addition to the three Oliver filters, the principal items of equipment in the concentrate filter plant are three 18 × 9 in. vacuum pumps, two 12 × 9 in. air compressors, two filtrate pumps and three mechanically agitated storage tanks for concentrate.

Labor requirements for concentrate handling are one operator and three Africans per shift and operating costs are 0.04s per ton milled, largely representing labor and power costs.

Tailing Disposal

Concentrator flotation tailing flows 2500 ft by gravity, to a 250-ft diam Dorr traction thickener, through two pipes in parallel, one of 18 in. and one of 22 in. diam. The clear thickener overflow passes directly to the main water storage tank and is returned to the mill circuit with enough make-up water to compensate for losses. Return water pumps consist of three of 1700 gpm and two of 850 gpm capacity, driven by 110 and 60 hp motors, respectively. Usually two large pumps and one small pump are sufficient to provide ample water at the concentrator at 35 psi. Water requirements are approximately 3 tons of water per ton of ore milled. Of this about 2 tons is return water from the thickener overflow, concentrate filtrate and others, and one ton is new make-up water obtained partly from water pumped from the mine and partly from the Luanshya stream.

Thickened tailing at 55 to 65 pct solids flows from the bottom of the thickener to a pump-surge tank and from there to the main tailing disposal pumps. Four 6-in. Wilfley pumps are available, each driven at 1286 rpm by 125 hp motors and each capable of delivering 4000 tons of solids per day through a 12-in. pipe line extending over 8000 ft to the furthest section of the tailing dam. An additional 8-in. Wilfley pump is kept as a stand-by for use in emergencies and only delivers to the nearest section of the dam. Discharge riser pipes, 6-in. diam and each with a valve, are spaced at about 300-ft intervals along the face of the dam.

The tailing dam is built along a gentle slope and as the dam increases in height there is also an increase in the area of the dam as it extends further up the slope. This calls for one main wall and two tapering wing walls that join with the natural slope of the terrain. A water basin forms where the dam meets the ground slope and the clear settled water is drawn off by diversion ditches provided with weirs that are raised to keep pace with increase in the general height of the dam.

Dam building is carried out by raising a wall of damp tailing some 12-in. high along the top edge of the dam, using material dug from a ditch running along the inside of the wall. This ditch then provides a means of distributing the tailing from the discharge pipe along the top of the dam wall. Tailing is run into every third section of the dam for one day and on the following day this section is cut off and the next section opened up. On the third day the last section is receiving tailing while the first section has drained sufficiently to enable the front of the wall to be built up again, ready for a further 3-day cycle. Each time that the outer wall is raised it is also pulled inward to give an average wall slope of 30° from the horizontal which has been found by experience to give the greatest all-round stability.

Although dam building is a relatively simple matter during the dry months, considerable difficulty is experienced in the wet season because of softening and sloughing of the outer face of the dam.

Tailing disposal labor consists of one operator on each shift with an additional charge-hand on day shift who is responsible for general supervision and for repairs to eroded sections of the dam. The African labor strength is 65 of whom 15 are divided equally among the 3 shifts and the remainder are engaged in building up the dam wall on day shift. Costs for tailing disposal are: labor, 0.03s; power, 0.03s; miscellaneous, 0.01s; total, 0.07 shillings per ton milled.

GENERAL

In conformity with modern design, the various concentrator buildings are of all-steel construction on heavy reinforced concrete foundations. Roofing and side covering consist of galvanized corrugated steel sheeting with ample provision for natural lighting. Reinforced concrete flooring is generally used for flooring in the crushing sections but in the milling and flotation sections the upper operating floors consist of wood decking on steel supporting string-

TABLE 3—*Operating Data*

CONCENTRATOR COSTS	COST PER TON MILLED
By Processes	
Crushing, conveying and screening	0.29
Grinding and classification	0.53
Flotation	0.45
Concentrate handling	0.04
Tailings disposal	0.07
Mill water	0.01
Assaying and sampling	0.02
General expenses	0.13
Total concentrating costs	1.54
	Kw-hr per Ton Milled
Power Consumption	
Crushing, conveying and screening	1.88
Grinding and classification	11.00
Flotation	3.26
Miscellaneous	1.62
Total power consumption	17.76
	Dry Tons per Man Shift
Labor Data	
European	180
Native	31
Total	27
	Avg. No. Shifts per Day
European	42
Native	241
Total	283

ers. Both wood and steel gratings have been tried for flooring but the former is unreliable and both have been found very tiring for the operating labor. Operating costs at the concentrator are given in Table 3.

Each main section of the concentrator (crushing plant, concentrator and filter plant) has its own sub-station to which electric power is transmitted from the power plant by underground cables at 3300 v, 3-phase, alternating current. Motors of above 125 hp are generally operated at 3300 v, lower power motors at 550 v and lighting at 110 v.

Milling Practice at Idarado Mining Company

By F. W. McQuiston, Jr.,* Member AIME

(Denver Meeting, October 1947, and New York Meeting, February 1948)

Introduction

Idarado Mining Company's mill and surface plant are at the portal of the Treasury tunnel at elevation 10,625 ft, 12 miles south of Ouray, Colo. In 1943 and 1944 this tunnel was extended in a westerly direction to tap the Black Bear vein, and, together with the recently completed 1100-ft inclined Treasury raise, provides access to the Black Bear mine. In 1945 a 250-ton mill was placed in operation to process development ores containing much needed copper, lead and zinc for war requirements, and to serve as a pilot testing unit for larger scale milling operations.

In the years 1902 to 1927 some 329,000 tons of ore were produced from mine workings on the Black Bear vein that extended downward from elevation 12,325 ft at the Black Bear tunnel portal to elevation 11,618 ft, the lowest level of the mine. The ores were taken to the Cascade and Smuggler-Union mills near Telluride over two aerial tramways. Gold bullion, a copper-lead concentrate, and a zinc concentrate were produced.

Ores

The Black Bear is a compound fissure vein that has undergone several stages of mineralization. The first important stage is represented by complex ores of copper, lead and zinc in a quartz gangue. Precious metal content in this stage is low. The galena is argentiferous and varying amounts of tetrahedrite, freibergite and bornite tend to increase the normal copper content represented in chalcopyrite. The zinc occurs as a true sphalerite with little or no iron. Later re-openings of the fissure took place more or less co-extensively with the successive introduction of (1) abundant quartz gangue containing minor amounts of chalcopyrite, galena and sphalerite; (2) free gold in a fluoritic quartz gangue; and (3) barren, low-grade quartz with minor amounts of pyrite as the only important sulphide. The sulphides show practically no oxidation and little mineralogic intergrowth. Because of the common overlapping of these successive stages of mineralization within a given mining area, considerable dilution of the base-metal content takes place, thus lowering the average mill-head grade.

Plant Practice

Metallurgical testing was started shortly after the Black Bear vein was encountered on Treasury tunnel level in 1944. Laboratory tests showed good flotability of the minerals in the ore and indicated an economic separation could be made of the copper, lead and zinc minerals into their respective concentrates. The success of the separation is close reagent control to maintain a balance between promotion and depression of the sulphides. To maintain this balance reagent consumption for the three metal separation is considerably higher than in ordinary lead-zinc flotation.

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A 65-mesh grind readily liberates the ore-bearing minerals from the gangue and from each other.

Plant grind is 6 pct plus 65 mesh and 50 pct minus 200 mesh. Grinding is done in closed circuit with a diaphragm-type jig and a spiral classifier. The free gold content of the ore makes it attractive to use a jig for removal of gold as bullion. Shipment of gold as bullion brings a higher price per unit and eliminates the hazards of sampling gold-enriched concentrates.

Grinding circuit, jig-hutch product is pumped to a cleaner jig. Cleaner jig-hutch product goes by gravity to a revolving continuous type amalgamator equipped with scrubber and amalgamation plates. Amalgamator tails are returned to the head of the rougher jig and cleaner jig tails returned to the ball-mill feed.

Amalgamator tailings assay as high in lead as the flotation lead concentrates; however, if taken out of the circuit the ratio of lead to copper would be lowered to a point that would upset the copper-lead float separation.

Jig practice employing continuous amalgamation effects a 60 pct gold recovery into bullion. Experimental work in process indicates a possibility of higher recoveries. The method now in use compares favorably with stamp battery and plate amalgamation.

The classifier overflow as feed to bulk copper-lead rougher flotation is maintained at 30 pct solids. After 10 min conditioning time the pulp is floated for 12 min in a 10-cell rougher machine. Scavenger concentrates from the last five rougher cells are returned to the head of the circuit. Rougher concentrates from the first five cells are cleaned and recleaned in single-cell machines. Cleaner tails flow countercurrent to the froth and are pumped back to the copper-lead conditioner. It has proved a desirable cleaning practice to float off a large volume of froth over a short length of cell lip to obtain a "crowding action." It has been established that best cleaning results are obtained when 8 to 10 tons of concentrates per 24 hr are overflowed per foot of froth weir. In addition high dilution is used advantageously in cleaning and recleaning.

Copper-lead cleaned concentrate is fed to the third cell of a 6-cell separation bank. A dilution of 9 to 1 is maintained in the circuit. Lead is floated in this circuit and froth overflow is returned from the last 3 cells to the third cell. Froth overflow from the third cell or feed cell goes to the second cell and thence to the first cell. The second and first cells are used as cleaner and recleaner, respectively, on the lead concentrate. Pulp underflow from the cleaner cells joins initial feed in the third cell. Pulp discharge from the sixth cell is the copper concentrate.

Copper-lead rougher tails is the feed to the zinc circuit. Launder spray water is controlled so that zinc flotation feed density is held at 25 pct solids. Feed to the zinc circuit is conditioned in two stages of 10 min each. The same flotation froth flow scheme is used in zinc roughing, cleaning and recleaning as in the copper-lead circuit.

Mine-water drainage is used for milling. The water enters the mill at an average temperature of 40°F. Milling results improve with higher pulp temperatures and steps are being taken to introduce heat to the pulps.

METALLURGY

The major metallurgical problem was promotion of chalcopyrite in the presence of zinc and iron depressants. Small quantities of cyanide were found to have a definite tendency to depress chalcopyrite. The sensitivity of chalcopyrite to cyanide proved to be a favorable feature in the separation of the galena from chalcopyrite and was the deciding factor in selecting cyanide in preference to the dichromate or sulphurous acid-zinc hydrosulphite methods of separation.

A 250 ton per day plant has been used for developing the metallurgical flow scheme worked out in the laboratory and for testing various types of milling equipment. The reagent setup developed in the laboratory has proved the most successful and is giving recoveries of each metal into separate concentrates greater than 85 pct.

Reagent practice is as follows: American Cyanamid reagent No. 404 is added to the ball mill along with lime, zinc sulphate, and cyanide. Reagent No. 404 was found to have the highest selectivity for galena of the many reagents tried. A slight excess has very little promotional effect on the pyrite, sphalerite and chalcopyrite in the presence of the depressants used. Lime serves a dual purpose: to give a pH of 9.5 at which the chalcopyrite floats best, and to retard flotation of the pyrite. With a lower pH the copper recovery falls off and the pyrite tends to float. Lime was selected for pH control in preference to soda ash. Soda ash, to some extent, is an activator of pyrite and also would carry through with the copper-lead tailings into the zinc circuit to combine with lime to form caustic soda which often has a tendency to encourage pyrite to float. Lime alkalinity at a pH of 9.5 has no depression effect on the free gold or galena. Zinc sulphate is added as a zinc depressant. Cyanide in minor amount is added as a zinc and pyrite depressant.

Amyl xanthate, which was found to be the most effective promoter of chalcopyrite, is added to the classifier pool. The xanthate is fed in such quantities that only the chalcopyrite requirements are cared for, as an excess pulls zinc and iron. A xanthate deficiency results in a copper loss into the tailings. If the chalcopyrite is over promoted it is difficult to depress satisfactorily in the copper-lead separation circuit. To allow for fluctuations of the copper content in the mill feed, the mill operators are allowed to add an additional 0.015 lb per ton. When the copper content of the ore exceeds the average, chalcopyrite can be observed floating sluggishly at the tail end of the rougher flotation circuit. The operators then increase the feeding rate of the xanthate and when pyrite has a tendency to float, the xanthate is decreased.

Du Pont frothing reagent B23 is added to the conditioner ahead of copper-lead rougher flotation. It is supplemented with additional B23 to flotation cells as required. Zinc sulphate and sodium sulphite, as zinc and pyrite depressants, are added to the bulk copper-lead flotation cleaner and recleaner.

In the separation circuit cyanide and sodium sulphite are stage fed directly to the flotation cells to depress the copper minerals so the galena may be floated. Conditioning with cyanide ahead of flotation proved detrimental to chalcopyrite depression. A possible reason is that conditioning with cyanide allowed time for the cyanide to dissolve copper forming a copper thiocyanate which reactivated the chalcopyrite. A trace of free cyanide is maintained in the separation circuit tailings to ensure sufficient but no excess cyanide being present during roughing and the several cleaning stages employed. From 5.0 to 8.0 lb of cyanide are required per ton of bulk copper-lead concentrate fed to the separation circuit. Sodium sulphite and cyanide have been found to be a good combination for depression of chalcopyrite. Best chalcopyrite depression is obtained when the pH is under 8.0, which is the reason the minimum amount of cyanide is used. It is definitely indicated that the copper-lead separation is better when the ratio of lead to copper is one and a half to one or higher.

When laboratory research was first started it did not seem practical for a mill separation to be made on copper-lead ores which contained less than one per cent copper; however, mill results definitely show an economic copper-lead separation can be made on ores containing one half of a per cent copper and possibly on ores with even a lower copper content. Effective separations have been made on bulk copper-lead concentrates assaying over 40 pct lead and less than 10 pct copper.

The tailings from the separation circuit become the copper concentrates and the flotation froth the lead concentrates. Activated carbon is agitated with the copper concentrates to adsorb the gold in solution dissolved by the cyanide in the separation circuit. The carbon containing gold goes with the copper for shipment to the smelter. A portion of the copper concentrates are tabled on a laboratory-size concentrating table which shows the amount of lead in the copper concentrates. The lead content in the copper concentrates is kept as low as possible even at the sacrifice of floating copper into the lead concentrates because of the more favorable smelter payment for copper in the lead concentrates over lead in the copper concentrates.

Tailings from copper-lead bulk flotation are conditioned in two stages; first with copper sulphate and butyl xanthate for activation and promotion, followed by secondary conditioning with lime for pyrite depression. Activation of the sphalerite with copper sulphate is more effective, at a low pH, and therefore, lime is added in secondary conditioning, after the sphalerite has been activated. Better zinc metallurgy would be obtained if the sphalerite could be activated in a neutral pH pulp; however, this cannot be practiced as lime is essential in copper-lead flotation. Heating the zinc pulp indicates improved metallurgical results in both zinc recovery and grade of concentrate. For zinc flotation a dry brittle froth is maintained by adding a slight excess of copper sulphate. Milk of lime is added to the zinc flotation cleaners to ensure pyrite depression. A pH reading is made hourly but is only used as an alkalinity indicator, as lime is added by predetermined quantities rather than by alkalinity control to make the desired zinc concentrate grade. Frothing is accomplished with B23.

Table 1 gives the amounts of reagents fed.

Table 1—*Amount of Reagents Used*

Reagent	Mill Feed, Lb per Ton	Place of Reagent Addition
A. C. No. 404.....	0.05	Ball mill.
Zinc sulphate......	1.50	Ball mill.
Sodium cyanide....	0.10	Ball mill.
Lime.............	2.00 (pH 9.5)	Ball mill.
Amyl xanthate....	0.025	Classifier pool.
Amyl xanthate....	0.015	Staged to copper-lead flotation.
Zinc sulphate......	0.20	Copper-lead cleaners.
Sodium sulphite...	0.20	Copper-lead cleaners.
Sodium cyanide....	0.30	Copper-lead separation.
Sodium sulphite...	0.15	Copper-lead separation.
Activated carbon..	0.25	Copper concentrate agitator.
Copper sulphate...	0.75	First zinc conditioner.
A. C. No. 301.....	0.10	First zinc conditioner.
Lime.............	6.00	Second zinc conditioner.
Du Pont B23......	0.20	Frother where required.

Metal recoveries for a month on float feed are given in Table 2.

Table 2—*Flotation Feed Recoveries*

	Percentage Distribution				
	Au	Ag	Cu	Pb	Zn
Flotation feed..	100.00	100.00	100.00	100.00	100.00
Copper concentrates.......	76.39	30.61	87.45	5.35	3.32
Lead concentrates.......	5.66	59.27	3.36	91.46	1.53
Zinc concentrates.......	4.90	4.68	5.73	1.79	88.14
Tailings.......	13.05	5.44	3.46	1.40	7.01

Plant

Metallurgical and mechanical testing has been sufficient to lay out a definite program for plant expansion.

Primary crushing will be done underground some 12,000 ft from the portal for several reasons. Surface conditions do not allow for economic installation of sufficient coarse ore storage. No additional buildings will be required over those now in existence. Construction underground is cheaper than surface buildings. No heating will be required. Carloading will be faster. Tramming costs will be lowered due to heavier loading per car with crushed rock over coarse shrinkage stope draw. Less close integration of mining and milling operations will be necessary.

Present surface crushing plant will be revised to allow a capacity of 100 tons per hour. Wooden rectangular surge bins will receive the ore ahead of secondary and tertiary crushing. Four reciprocating feeders discharge from the bins on to a conveyor for feeding a 3-ft reduction crusher which will deliver a 2-in. product. The feeders will be electrically controlled from the crusher operating floor. Secondary crusher discharge will be conveyed to a 4 by 8-ft rod deck screen operating in closed circuit with a 4-ft cone crusher. Both crushers will be protected from mine iron by magnets. A minus ⅜-in. screen undersize is elevated to a 1000 ton fine ore bin.

The fine crushing plant is designed to be operated by one man as both the secondary and tertiary crusher will be adjacent to each other. An electrical interlocking system shuts off all motors in the event any single motor cuts out, and a simple light system indicates which one. Both the underground and surface crushing plants will be equipped with 24-in. conveyors. Each crusher discharge and all conveyor places will be hooded for dust collection. A multiclone dust collector removes the dust from the collector lines.

Large storage bin capacity has been provided to ensure thoroughly mixing the ore to give a uniform mill feed which is an important factor in metallurgical results. Fine ore will feed from three chutes to a belt driven by a variable speed motor. Belt discharge will be to the ball-mill feed belt which will be equipped with a recording weightometer. Automatic grinding-circuit control will be affected through the first belt.

A 7 by 7-ft grate mill is installed to operate in closed circuit with a jig and 60-in. spiral classifier. The grinding and classifying unit are estimated to have a 450 ton daily capacity. Flotation capacity of 500 tons per day will be provided.

Experimental work was conducted with various types of flotation machines run at different speeds. Rotor and impeller wear was correlated with metallurgical results. Standard impeller peripherial speeds of 1600 fpm made fair metallurgical recoveries; however, increasing the peripherial speed to 1900 fpm gave much better aeration with a smaller bubble and livelier froth which yielded lower tailings. Better aeration reflected in lower copper-lead and zinc rougher tailings, but was most pronounced in giving a better copper recovery.

The Black Bear ore is quite abrasive and with an impeller speed of 1900 fpm wear took place quite rapidly. Iron impellers lasted only a few weeks and pre-war rubber-covered impellers wore badly in a few months. The impeller wear reflected directly in a higher tailing. Increasing the peripherial speed to approximately 2200 fpm, to compensate for wear, raised the recovery but also greatly increased impeller wearing. Plant experimental work showed flotation machines designed with vertical impeller blades gave satisfactory metallurgical results on Black Bear ore when the impeller traveled approximately 1900 fpm. To maintain results, however, wear had to be compensated by increasing peripherial speeds. Due to speed of wear this was impractical from a cost and operating standpoint.

Rotor-type flotation machines run at high speeds gave satisfactory results and a minimum reflection in tailings losses with rotor and stator wear. Iron rotors and stators gave a short life when run at 2150 fpm peripherial speed. Natural rubber rotors and stators, however, give satisfactory wearing life at high speeds.

On a 20 cu ft cell, with partially worn impellers, running at 1600 fpm, each cell motor pulled approximately 1.5 hp. Increasing the speed to 2250 fpm doubled the horsepower requirements; however, the copper tailings dropped 0.04 pct copper. When power costs, with normal impeller speeds, are compared with the cost of power for high speed operation the recovery of more pounds metal favors the use of extra power.

Conclusions were that rotor-stator type machines at higher than standard speeds would be used on all scavenger rougher circuits. Impeller-type machines at standard speeds would be used on all cleaner circuits.

For copper-lead roughing 40 cu ft rotor-type cells will be used. On a 10-cell machine the rotors in the first two cells will run 1600 fpm and pull just under 5 hp per cell. The next two cells will run 1800 fpm with 6 hp per cell, and the six scavenger cells will run 2150 fpm and use 8 hp per cell. Similar speeds will also be used in the zinc-rougher flotation bank.

Impeller-type machines operated at normal speed are used on the copper-lead cleaner and recleaner circuits. This type of flotation machine is used because of the advantage in introducing flotation products into any cell without the aid of pumps.

The practice of using high-speed rotor-type flotation for intense agitation and aeration in roughing combined with slow-speed impeller agitation for cleaning has given good metallurgical results.

The flotation operating floor is on one level to facilitate efficient control. Flotation circuits for copper-lead, and zinc roughing, cleaning, recleaning and separation will be operated from this one level. A total of three pumps will be required for the entire flotation circuits.

No sunlight will be admitted through windows and illumination will be entirely with fluorescent lighting. Sunlight reflected from the snowbanks into the mill causes discomfort to the operators and also changes the color appearance of the flotation froth.

Concentrate thickeners are designed to require a minimum of floor space. A single tank with two compartments and one drive mechanism will be used. The design is such that there is no possibility of leakage from the top to bottom tray. Two such units will be required for contemplated tonnage.

Thickened concentrates will be filtered on disc filters for lead and copper and a drum filter for zinc. Filter discharge will drop into open-faced bins. Truck loading of concentrates will be by use of a double-drum slusher hoist scraping from the bins directly into trucks.

SUMMARY

Idarado Mining Company is recovering five metals—gold, silver, copper, lead and zinc. Recovery is affected in four products. Gold is shipped as bullion to the U. S. Mint, Denver; concentrates are trucked to Montrose, Colo. and thence by rail to the smelters. A copper concentrate, a lead concentrate and a zinc concentrate are shipped to respective smelters. Each concentrate produced carries values in gold and silver.

Metallurgy at National Lead Company, MacIntyre Development

By Frank R. Milliken,* Member AIME

(New York Meeting, February 1948)

Scope

This paper is a running commentary on metallurgical problems and developments, stressing ilmenite flotation, since the start of operations five years ago, at the mill of National Lead Company, Titanium Division, MacIntyre Development, Tahawus, New York.

Table 1—*Ore Types*

Type Number	Local Name	Approximate Mineralogical Composition				Economic Grind Required for Liberation, Mesh	
		Magnetite	Ilmenite	Feldspar	Iron Silicates	Magnetite from Ilmenite	Magnetite and Ilmenite from Gangue
1	High grade anorthosite ore	47	38	13	2	48	20
2	Low grade anorthosite ore	35	25	35	5	48	20
3	High grade coarse-grained gabbro ore	40	40	5	15	48	20
4	High grade fine-grained gabbro ore	40	40	5	15	48	65
5	Disseminated or low grade gabbro ore	25	25	20	30	65	65
	Average	37	32	16	15		

General

In early 1941, when world conditions threatened to cut off the supply of ilmenite (obtained largely from India) to this country, the National Lead Company started to develop and bring into production their MacIntyre Development. This operation supplied the bulk of the ilmenite used during the war period for the manufacture of the much needed titanium pigments.

MacIntyre Development is at Tahawus, New York, on the headwaters of the Hudson River in the center of the scenic Adirondack Mountains. A general description of the operation was given in the Adirondack Issue of Mining and Metallurgy for November 1943. Operating tonnages and production have increased since 1943, however. Nearly 10,000 tons of ore and waste is broken and loaded daily in the open-pit mine. About half of this tonnage is ore, from which is produced nearly 800 tons of ilmenite concentrate and 1600 tons of magnetite concentrate daily.

Ore—types

From an ore-dressing standpoint, there are five rather distinct ore types and two waste-rock types in the mine. The ore types are shown in Table 1.

The anorthosite waste is a coarse-grained white rock, largely labradorite feldspar. The gabbro waste is generally finer grained, and contains, in addition to feldspar, 35 to 75 pct iron silicates such as hornblende, pyroxene, garnet and biotite. A minor mineral mineralogically, but important metallurgically, is apatite. A few calcite

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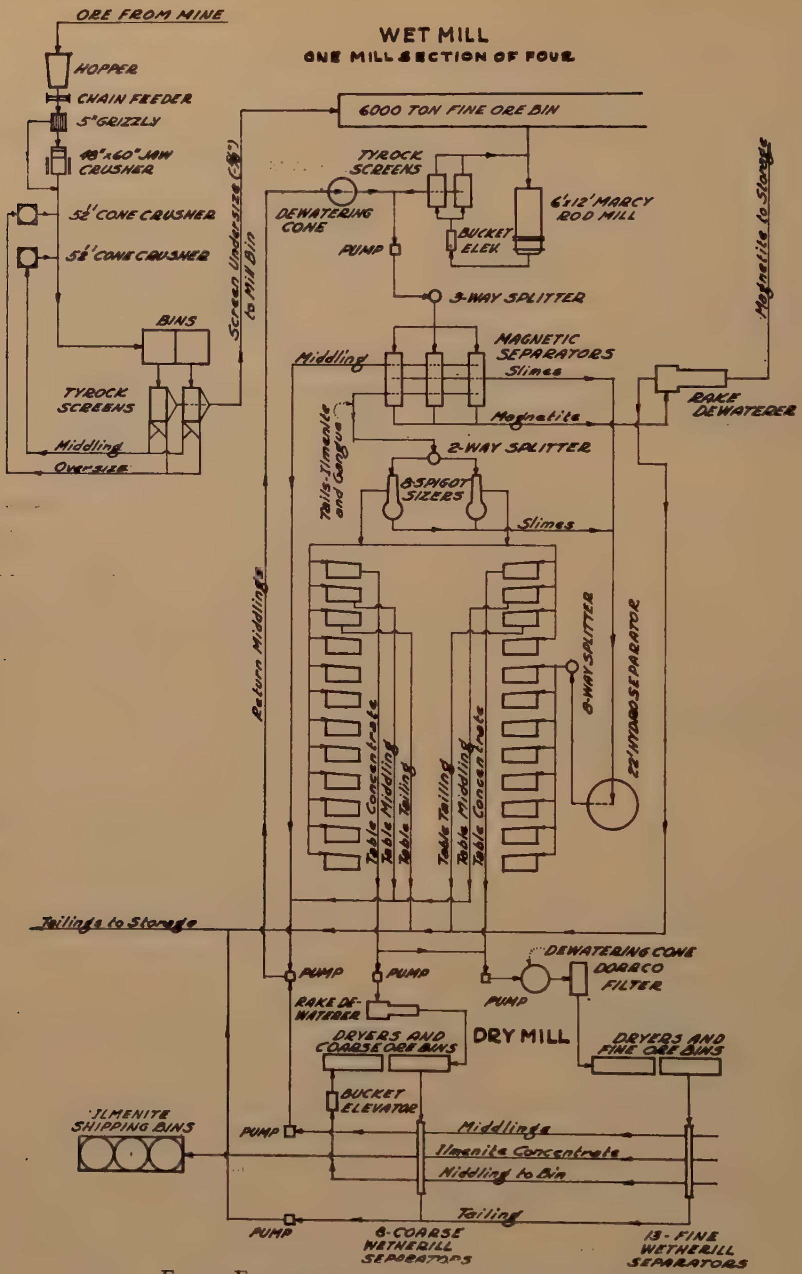

FIG 1—FLOWSHEET OF MILLING OPERATIONS, 1942.

stringers occur at isolated locations in the ore body.

The ore and waste are very hard and tough (resistant to both crushing and grinding) and a minimum of primary slimes and grinding slimes occur.

ORE—TESTING

The original MacIntyre mill design was based on testing conducted over a period of years by Professor Charles E. Locke of the Massachusetts Institute of Technology with whom in the earliest tests was associated the late Professor Robert H. Richards. Samples for testing were all of type No. 1 (high grade anorthosite ore).

Tests on this ore type showed:

1. A 20-mesh grind was adequate to liberate the gangue minerals from the magnetite and ilmenite.

2. At a 20-mesh grind, a magnetite product could be recovered by magnetic separation, leaving an ilmenite-gangue tailing, but this magnetic product assayed 10 pct TiO_2 and contained about 30 pct of the ilmenite in the crude ore.

3. Gravity concentration of the ilmenite-gangue tailing from the magnetite separation gave an ilmenite product assaying about 44 pct TiO_2 with a loss of 10 to 15 pct of the ilmenite in the crude ore in table tailing and (more largely) in slime products. The ilmenite gravity concentrate contained some gangue minerals.

4. Wetherill magnetic separation of the ilmenite gravity concentrate eliminated additional gangue, raising the ilmenite grade to 48 pct TiO_2 with a small loss of ilmenite. The expense of the Wetherill separation and small additional TiO_2 loss were economically justified by savings in subsequent processing costs of the ilmenite at pigment plants.

ORIGINAL MILL FLOWSHEET

From the above test results, the original mill flowsheet, shown in Fig 1, was evolved. Briefly, the ore was crushed from run-of-mine size (some pieces up to 5 ft in one dimension) to minus-¾ in. square mesh in three stages, using a jaw crusher and cone crushers.

The crushing plant product was then ground in open-end rod mills in closed circuit with mechanically vibrated screens to approximately minus-20 mesh. The minus-20 mesh screen undersize was treated in Crockett magnetic separators which produced magnetite concentrate, nonmagnetic tailings (containing the ilmenite and gangue minerals), a middling (returned) and a slime overflow. After dewatering, the magnetite went to direct shipment, a sintering plant, or stockpile. The ilmenite-gangue mixture was then classified to make eight separate sand spigot products and a slime overflow. The spigot products were tabled separately. The combined classifier and Crockett slime overflows were thickened and tabled. Tabling operations produced a rough ilmenite concentrate and a tailing, middlings being returned as shown. The rough table concentrate was dewatered and dried, then treated on Wetherill magnetic separators. These Wetherills produced final ilmenite concentrate, two returned middling products, and another tailing product.

PROBLEMS ENCOUNTERED

Early mine development soon revealed, however, that only about 25 pct of the ore reserves were of type No. 1. At least 30 pct were of type No. 5, with about 15 pct each of types No. 2, 3 and 4. Thus, instead of an ore containing 47 pct magnetite, 38 pct ilmenite, 13 pct feldspar, and 2 pct iron silicates, the average mill feed was actually about 37 pct magnetite, 32 pct ilmenite, 16 pct feldspar, and 15 pct iron silicates. These differences do not seem particularly significant except, perhaps, for the lower tenor of the ore; however, from a metallurgical standpoint, these changes had great effect. The important factors were:

1. The decreased mill-feed magnetite content decreased magnetite-concentrate production from about 60 pct by weight of the mill feed to about 46 pct by weight of the feed, or increased the Crockett tailing tonnage from 40 pct of the feed to 54 pct of the feed. This meant an increase in classifier and table-feed tonnage rates of about 35 pct.

Table 2—*Metallurgical Balances*[a]

Product	Tons	Weight, Pct	Assay, Pct		Content, Tons TiO_2	Distribution, TiO_2, Pct
			TiO_2	Fe		
Case 1—Original Tests, Type No. 1 Ore						
Magnetite	1,998	60.0	10.0	58.0	199.8	30.0
Ilmenite	798	24.0	48.0		383.0	57.5
Tailing	534	16.0	15.6		83.2	12.5
Heading	3,330	100.0	20.0		666.0	100.0
Case 2—Former Results, Average Ore						
Magnetite	1,576	47.3	10.0	56.5	157.6	28.0
Ilmenite	614	18.4	44.0		270.1	48.0
Tailing	1,140	34.3	11.9		135.1	24.0
Heading	3,330	100.0	16.9		562.8	100.0
Case 3—Present Results, Average Ore						
Magnetite	1,632	49.0	10.0	58.3	163.2	26.5
Ilmenite	763	22.9	44.8		341.9	55.5
Tailing	935	28.1	11.9		110.9	18.0
Heading	3,330	100.0	18.5		616.0	100.0

[a] Based on 3330 tons of mill feed daily. Mechanical improvements have permitted an increase in daily tonnage over 3330 tons per day so that actual production at present is close to 800 tons of ilmenite per day.

2. The decreased quantity of ilmenite in the mill feed and the increased quantity of Crockett tailing, decreased the ilmenite content of the classifier and table feed. Whereas the gravity-section feed was expected to average 35 pct TiO_2, it actually averaged 22 pct TiO_2. The increase in gravity section feed rate tonnage from about 1300 to 1800 tons per day and the decrease in the gravity section feed ilmenite grade (from 35 pct TiO_2 to 22 pct TiO_2) had an adverse effect on table metallurgical results so that both ilmenite-concentrate grade and ilmenite recovery were lower than desired.

3. The proportion of iron silicate minerals to ilmenite was increased from 1:19.0 up to 1:2.1. These iron silicates were materially higher in specific gravity than feldspar; some of the minerals, particularly garnet, were close to ilmenite in specific gravity (3.8 to 4.0 for garnet and 4.4 to 4.6 for ilmenite). This made the separation of ilmenite from gangue minerals by gravity much more difficult than was expected. In addition, these iron silicates showed a magnetic susceptibility approaching that of ilmenite, and therefore could not be rejected in Wetherill operations without excessive ilmenite losses. Combined tabling and Wetherill operations on the so-called plant slimes (thickened slime overflows from the Crockett separators and the classifiers) were particularly ineffective.

4. The presence of a large proportion of disseminated ore, in which the ilmenite was not liberated from the gangue at the 20-mesh grind employed, further adversely affected metallurgical results.

The practical effect of the above differences was to decrease both ilmenite production per day and ilmenite grade. Where, from 3330 tons of mill feed per day, a production of nearly 800 tons of 48 pct TiO_2 ilmenite had been anticipated, actual production was only 614 tons per day of 44 pct TiO_2 ilmenite.

1. Increasing mill-feed grade to permit increased production, both through higher feed metal content and improved percentage recovery.
2. Grinding finer to provide adequate liberation.

TABLE 3—*Metallurgical Balance, Cobbing Operation*

Product	Tons	Weight, Pct	Assay, Pct TiO_2	Content, Tons TiO_2	Distribution, TiO_2, Pct
Concentrate (to mill)	3,330	84.7	18.7	623	96.0
Reject (to waste)	600	15.3	4.3	26	4.0
Feed (from mine)	3,930	100.0	16.5	649	100.0

TABLE 4—*Assay—Screen Analysis*

Size, Mesh	Magnetite Concentrate, Pct				Tailing, Pct		
	Weight, Direct	Assay		Distribution, TiO_2	Weight, Direct	Assay, TiO_2	Distribution, TiO_2
		TiO_2	Fe				
+ 28	17.5	11.8	54.2	20.2	2.9	6.7	1.5
−28 + 35	15.1	11.6	54.2	17.2	4.5	6.7	2.4
−35 + 48	14.8	10.5	56.5	15.2	7.9	7.1	4.5
−48 + 65	13.6	9.4	58.1	12.6	9.7	7.3	5.8
−65 + 100	9.9	9.4	57.9	9.1	11.9	9.7	9.2
−100 + 150	7.4	8.6	57.9	6.3	10.7	9.1	7.8
−150 + 200	6.8	8.6	56.1	5.8	11.6	12.6	11.7
− 200	14.9	9.3	56.1	13.6	40.8	17.6	57.1
Total	100.0	10.2	56.1	100.0	100.0	12.5	100.0

The relative metallurgical balances for these cases (Case 1 and Case 2) are given in Table 2.

TABLE 5—*TiO_2 Recovery by Sizes*

Size, Mesh	Percentage TiO_2 Recovery by Size, Approximate		
	Magnetite	Ilmenite	Tailing
+ 28	55	37	8
− 28 + 35	35	53	12
− 35 + 48	28	59	13
− 48 + 65	23	62	15
− 65 + 100	22	62	16
− 100 + 150	22	60	18
− 150 + 200	23	53	24
− 200	25	16	59

It was evident that some plant changes would be required to improve metallurgical results and production. The general problems appeared to be:

3. Improving gravity concentration efficiency to partially compensate for the more difficult separation involved.
4. Developing a concentration procedure more adapted to the required separation of ilmenite from gangue, particularly for the finer sizes.

It was possible to effect a material increase in mill-feed grade by installing a magnetic cobbing operation between the second and third stages of crushing. This cobbing operation (treating two sizes: plus 1¾ in. and minus 1¾ in. plus ⅝ in.) rejects a nonmagnetic tailing product containing very little ilmenite and magnetite. A typical metallurgical balance of cobbing operations is given in Table 3.

The grinding problem is best illustrated by assay-screen analyses. Table 4 shows the assay-screen analysis of May 1944, monthly composite samples, and Table 5 shows TiO_2 recovery by size in the three mill products (magnetite, ilmenite, and tailing). Table 6 shows the effect on liberation of regrinding plant magnetite concentrate. The reground concentrate was treated in a Davis tube tester to produce a cleaned magnetite concentrate and a tailing of ilmenite and gangue minerals liberated by the regrinding. The screen analysis of the magnetite concentrate (Table 4), the high loss of TiO_2 in the coarse sizes of the magnetite, and the liberation of ilmenite by further grinding of the magnetite (note in Table 6 that by regrinding to 65 mesh more than one-third of the TiO_2 lost in the plant magnetite concentrate is liberated), all showed the necessity for finer grinding for adequate liberation. The high loss of minus-200 mesh ilmenite in the plant tailings (note in Table 4 that the minus-200 mesh tailings assayed nearly 18 pct TiO_2 and in Table 5 that 59 pct of the minus-200 mesh ilmenite in the feed was lost in the tailings), however, argued against finer grinding unless some suitable concentration process for recovery of fine ilmenite could be developed. A further limitation on finer grinding was the adverse effect of the finer magnetite on the subsequent magnetite sintering operations.

TABLE 6—*Davis Tube Tests on Magnetite Concentrate*[a] *Reground to Varying Degrees of Fineness*

Size Reground to	Davis Tube Concentrate, Pct				Davis Tube Tailing, Pct		
	Weight	Assay		TiO_2 Distribution	Weight	Assay, TiO_2	TiO_2 Distribution
		TiO_2	Fe				
All through 28 mesh	92.8	8.8	59.0	80.1	7.2	28.1	19.9
All through 35 mesh	89.8	8.5	59.6	74.8	10.2	25.2	25.2
All through 48 mesh	86.6	7.9	61.3	67.0	13.4	25.1	33.0
All through 65 mesh	84.8	7.6	61.5	63.1	15.2	24.7	36.9
All through 100 mesh	84.2	7.3	61.2	60.2	15.8	25.6	39.8

[a] Assaying 10.2 pct TiO_2 and 56.7 pct Fe.

To improve gravity concentration metallurgy, it was determined that better classification and additional tables were needed. Classification was improved by replacing the two original overloaded eight pocket Fahrenwald sizers in each section with one phosphate rock-type Dorrco sizer of adequate capacity with electrical controls for maintaining pocket density. The installation of additional tables was an unsolved problem, however, because of space limitations.

FLOTATION

Soon after plant operations started, flotation testing for ilmenite recovery was initiated as the obvious method for recovering the fine ilmenite lost in the gravity concentration and dry magnetic separation tailings. In addition to reducing fine ilmenite losses, it was expected (1) that a successful flotation procedure would permit a finer original grind (within the limits dictated by magnetite sintering) without increased losses of fine ilmenite and (2) that slime table feed (thickened Crockett separator and classifier slimes) could be treated in a flotation plant requiring little floor space and that the tables so released from service could be used on classifier sands. Thus, flotation provided partial answers, at least to the general problems 2, 3 and 4 involved in improved metallurgy, as stated earlier. On a long range basis, it was also felt that a satisfactory flotation procedure might ultimately permit an original 65-mesh grind, required for adequate liberation, with complete replacement of the gravity concentration and dry magnetic separation sections by flotation for ilmenite recovery.

Testing was concentrated on the slime table feed, as the greatest immediate benefits would result from improved metallurgy on that product. This thinking must parallel the early work on flotation of sulphide ores, when the necessity for reducing the slime losses of gravity concentration was paramount. In the initial work, practically all known promoters were investigated. In general, cationic reagents (amines, quaternary ammonium compounds, pyridinium salts) showed little selectivity between ilmenite and the gangue minerals (particularly the iron silicates). Soaps, resinates, the lower fatty acids, tall oils, and napthenic acid, although all showing some collecting properties for ilmenite, gave poor recoveries with partial selectivity. All these collectors were tested over wide pH ranges and in combination with various modifying agents. Initial tests with oleic acid, possibly the universal nonmetallic collector, showed immediate promise, however. Although initial tests, conducted at pH 7.2 to 7.5, gave rather low grade concentrates (40 to 42 pct TiO_2), recoveries were satisfactory.

Further work on this simple anionic float, investigating conditioning time and density, collector amounts, flotation density, and pH, evolved a test procedure in which the slime table feed was floated directly with 0.50 to 0.60 lb of oleic acid per ton of feed and sufficient sulphuric acid to give a pH of 7.2. The ilmenite rougher concentrate was cleaned twice without additional reagents. When the plant was treating an ore mixture high in type No. 1 ore, laboratory flotation tests on samples of slime table feed were fairly satisfactory (on a slime table-feed sample assaying 20.0 pct TiO_2, a recovery of 88 pct of the TiO_2 in a concentrate assaying 45.0 pct TiO_2 was obtained). Using this identical procedure when the mill feed was high in type 5 ore, results were less satisfactory (on a slime table-feed sample assaying 19.0 pct TiO_2, a recovery of 85 pct of the TiO_2 in a concentrate assaying 43.0 pct TiO_2 was obtained); the iron silicate minerals, particularly garnet, were found to float readily with oleic acid.

TABLE 7—*Comparison of Neutral Circuit Flotation with Acid-wash Flotation Circuit*

Type Ore	Preponderantly Type No. 1		Preponderantly Type No. 5	
Circuit	Neutral	Acid-wash	Neutral	Acid-wash
Heading, TiO_2, pct....	23.7	23.4	19.5	19.3
Final concentrate, TiO_2, pct..........	44.9	46.8	42.2	45.1
TiO_2 recovery in final concentrate, pct.....	90.4	90.4	84.6	85.2
Tailing, TiO_2, pct.....	4.5	4.2	4.9	4.5
Oleic acid, pound per ton feed...........	0.47	0.50	0.45	0.50
Sulphuric acid, pound per ton feed........	0.40	9.51	0.50	10.05

As there were several literature references regarding low pH circuits for metallic oxide flotation, considerable testing with low pH circuits was done. With a low pH in the rougher circuit, however, collector consumption was excessive. The use of a neutral rougher circuit followed by cleaning in low pH circuits was then tried. Finally, conditioning of the rougher concentrate at a low pH, followed by washing of the acid-conditioned rougher concentrate, then reflotation, was developed. The iron silicates which floated in the rougher circuit were effectively depressed in the reflotation step following acid conditioning and washing. Comparison of this latter procedure with the former neutral circuit flotation is given in Table 7.

As results with the acid-wash procedure appeared satisfactory on all ore types, it was decided to build a pilot plant to test both circuits. The pilot plant was designed to handle the slime table feed from two of the four mill sections (about 150 to 200 tons per day). Principal equipment consisted of a 6 by 6 ft primary conditioner, two 6-cell No. 24 Denver flotation machines (6 cells for roughing, 3 for cleaning, and 3 for recleaning), two 5 by 5 ft conditioners for use with the acid-wash circuit (wood tank, rubber-covered mechanism), one 18-ft diam concentrate thickener, a four disc 6-ft diam disc filter, and auxiliary pumps and reagent feeders.

The pilot plant was started up on the simple neutral-circuit flowsheet. Metallurgical results were below expectations from laboratory work but shipments of satisfactory grade ilmenite flotation concentrate were made to the pigment plants using MacIntyre ilmenite. Trouble was immediately encountered at the pigment plants. This trouble, not noticed in preliminary pigment tests on a laboratory scale, was finally traced to the presence of about 1.0 pct P_2O_5 in the ilmenite flotation concentrate. Minor amounts of apatite in the crude ore, rejected by tabling and magnetic methods, were being concentrated with the ilmenite in the flotation product. Testing quickly determined that although the P_2O_5 was lower in concentrate produced by the acid-wash procedure, it was still considerably above the 0.10 pct P_2O_5 limit set by the consumers.

The pilot plant was therefore shut down and fundamental testing was again started. It was then found that the low pH rougher circuits, particularly with fluoride ion present, formerly investigated, rejected apatite. However, reagent costs of those circuits were excessive. It was next determined that reagent costs could be reduced materially by the use of some fuel oil. The fuel oil reduced oleic acid consumption to a reasonable figure, and with the reduced oleic acid, permitted a higher pH and reduced quantities of fluoride salt without any decrease in selectivity. The final reagent combination developed in the laboratory therefore comprised about 1.0 lb oleic acid, 1.4 lb fuel oil, 0.8 lb sodium fluoride, and sufficient sulphuric acid to give a conditioning pH of 6.2. Conditioning was for 10 min at 40 pct solids; all reagents were added to conditioning. Rougher flotation was at 20 pct solids; the rougher concentrate was cleaned twice with small sulphuric acid additions to hold cleaner pH at 5.5 to 6.0. The final ilmenite concentrate produced on average ore assayed about 46.5 pct TiO_2 and only a trace (less than 0.02 pct) P_2O_5. Recovery was 88 pct.

The two acid-wash conditioners were relocated in the pilot plant, and pilot-plant operation was resumed, duplicating as nearly as possible the laboratory conditions just outlined. Initial plant results were disappointing. Ilmenite grade was about 45.0 pct TiO_2 and recovery was 70 pct. Laboratory results on samples cut from the pilot-plant feed were very satisfactory, however. It was then determined that when the plant-conditioner discharge was floated in the laboratory (all reagents had been added to conditioning), laboratory results were relatively poor, being similar to plant-flotation results. The discrepancy between laboratory and plant was therefore evidently in conditioning. Contact times, aeration, densities and temperature were carefully checked, then a series of plant investigations were made adding reagents in different sequences, but no improvements resulted. It was finally noticed in the laboratory that tests conditioned at low conditioner impeller speeds appeared different from standard tests. This observation led to the series of tests shown in Tables 8 and 9.

These comparative tests on slime table feed showed that at low impeller speeds, selectivity between ilmenite and gangue minerals was poor, even with longer conditioner contact times. Laboratory flotation results on pilot-plant conditioner discharge were similar to results obtained at a laboratory conditioner impeller speed of 900 to 1000 rpm.

Two investigations were then started, the first in the plant, attempting to duplicate laboratory results in the plant conditioners, and the second in the laboratory to determine the reason for the improved results at higher impeller speeds.

TABLE 8—*Effect of Laboratory Conditioner Impeller Speed*[a]

Laboratory Impeller, rpm	Test Heading, TiO_2, Pct	Rougher Concentrate, Pct		Cleaner Concentrate, Pct		
		TiO_2	TiO_2 Recovery	TiO_2	P_2O_5	TiO_2 Recovery
500	16.1	28.5	96.0	34.2	0.098	90.3
900	16.2	29.1	95.7	34.5	0.148	90.0
1,200	16.5	33.5	97.5	39.7	0.061	92.5
1,500	17.2	37.4	90.9	42.9	0.015	86.2
1,900	17.1	38.0	93.2	43.3	Trace	88.7
Plant conditioner discharge	16.8	28.6	95.0	36.0	0.091	88.5

[a] Test constants: 10 min conditioning at 40 pct solids with 1.0 lb oleic acid, 1.4 lb fuel oil, 0.85 lb sodium fluoride, and 2.4 lb sulphuric acid per ton feed. 7 min rougher flotation at 20 pct solids.

TABLE 9—*Effect of Laboratory Conditioner Impeller Speed and Contact Time*[a]

Conditioner Time, Min	Impeller, rpm	Test Heading, TiO_2, Pct	Rougher Concentrate, Pct		Cleaner Concentrate, Pct	
			TiO_2	TiO_2 Recovery	TiO_2	TiO_2 Recovery
5	900	25.5	42.4	63.6	45.6	48.6
20	900	26.2	40.6	81.3	45.5	68.8
5	1500	25.4	41.8	87.5	46.2	71.4
10	1500	21.8	39.6	92.5	44.7	83.0
15	1500	25.0	41.5	94.8	45.6	89.0
20	1500	23.1	40.4	96.5	45.2	89.5
5	1900	23.8	42.9	86.5	46.8	72.5
10	1900	23.1	42.2	93.1	46.4	86.7
15	1900	23.2	42.9	94.0	45.2	85.6
20	1900	22.7	43.6	94.1	46.6	88.5

[a] Test constants: condition at 40 pct solids with 1.0 lb oleic acid, 1.4 lb fuel oil, 0.85 lb sodium fluoride, and 2.4 lb sulphuric acid per ton feed. Rougher flotation for 7 min at 20 pct solids.

In the plant, the following steps were taken:

1. The plant conditioners were 5 ft in diameter by 4 ft deep. Each was equipped with a 10 hp motor and a 24-in. diam rubber-covered ship-type impeller. Center wells were employed. Impeller speed was first increased to the maximum permitted by available motor horsepower with little improvement.

2. Next 15 hp, then 20 hp motors were installed and various methods of baffling to increase violence of agitation with and without center wells and at various speeds were tried. A visual comparison of plant and laboratory agitation indicated the plant units were providing more intense agitation than the laboratory unit, but plant and laboratory flotation results on the plant conditioner discharge showed only moderate improvement.

3. A turbo-mixer type impeller was tried, but results were inferior.

4. Standard parts for a Denver No. 30 flotation machine (receded disc impeller and diffuser plate) were installed, with a central standpipe to provide maximum circulation. Metallurgical results improved immediately; plant results were only slightly inferior to laboratory tests.

In the meantime, laboratory investigations of conditioning were continued. It was thought pre-emulsification of reagents, either mechanical or chemical, might permit less violent conditioning, but emulsification had no effect, nor did reagent saponification. The possibility that attrition scrubbing might be taking place was also considered, but no amount of violent conditioning prior to reagent addition was of benefit. Preconditioning at high intensities with any of the reagents other than oleic acid and fuel oil also had no effect. Final practical conclusions were that intense agitation with oleic acid and fuel oil using a shearing-type impeller with maximum circulation through the impeller was

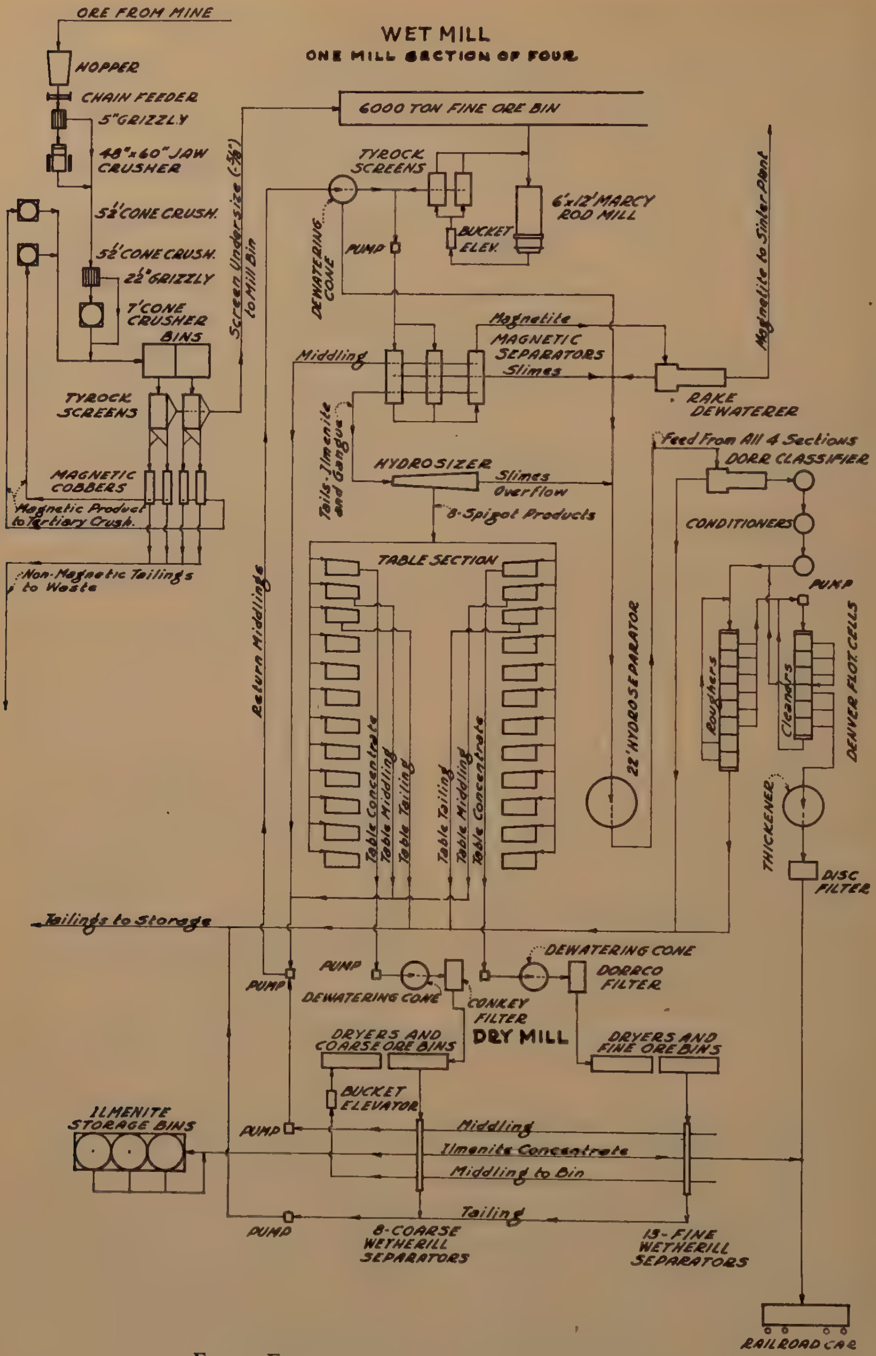

FIG 2—FLOWSHEET OF MILLING OPERATIONS, 1947.

required for good ilmenite recovery and gangue rejection.

Although laboratory testing had indicated no metallurgical benefits from conditioning at higher densities, neither were results inferior. Higher density conditioning had not been tried in the plant, as 40 pct solids was the thickest pulp which could be delivered to the conditioners from the sand pumps elevating the thickened slime table feed to the pilot plant. It was decided to install a rake classifier ahead of the pilot plant. It was felt that this classifier would serve three purposes:

1. Although the slime table feed thickeners served to some extent as deslimers, the classifier would eliminate practically all colloidal slimes from the flotation feed.

2. The classifier would even out feed-pump surges and an accurate feed tonnage determination could be made by catching the sands from one rake stroke.

3. A higher density in both conditioning and flotation could be obtained, so that plant capacity would be greater.

A further improvement in plant results occurred when the higher density conditioning was tried, so that plant and laboratory results were finally substantially identical. With the high density conditioning, it was also found that a simple impeller with four radial blades set at 45°, operated as a downthrust unit, gave satisfactory results. This impeller was 24 in. in diameter, and operated at 260 rpm.

At the higher density, plant capacity was adequate to handle the former slime table feed from all four mill sections, so that the slime tables were released for service on classifier sands. With flotation on all the slime table feed, it was also possible to grind slightly finer (to about 4 pct on 28 mesh compared to 10 to 12 pct plus-28 mesh), but this was still limited by sintering plant demands.

PRESENT OPERATIONS

The present flotation flowsheet, and the position of flotation in the present general plant flowsheet is shown in Fig 2. The feed to the flotation feed dewatering classifier consists of the underflow from the four 22-ft diam hydroseparators handling the slime overflows from the Dorrco sizers, the Crockett separators, the middling cones, and the magnetite dewaterers (tonnage from the latter two sources is minor). The Dorr rake classifier sands at 63 pct solids drop into the first of three conditioners in series which provide about 18 min contact on the average feed rate of 270 tons per day Oleic acid, fuel oil, sodium fluoride, and sulphuric acid (1.0, 1.4, 0.5, and 2.0 lb, respectively) are fed to the first conditioner. Sulphuric acid is regulated to give a pH of 6.5 in the first conditioner and small additions are made to the second and third conditioners to hold the pH at 6.5. This has been found to be important. After conditioning, a rougher concentrate is produced in eight Denver No. 24 cells (sometimes the concentrate from the last two or three cells is returned to the head of the rougher circuit). The rougher concentrate is cleaned in three cells and recleaned in three cells of the same type except that both cleaner and recleaner cells are of wood construction while the rougher is steel construction. Acid is added to the cleaner to give a pH of 5.8 and to the recleaner to give a pH of 5.4. Cleaner tails return to the rougher and recleaner tails to the cleaner. A two-cell Fagergren machine and a two-cell Agitair machine have, at times, been used in the rougher circuit for comparative testing.

Final concentrate is thickened, after the addition of lime, in an 18-ft diam thickener, and is then filtered in a four disc 6-ft diam filter. Filter cake, at 10 to 11 pct moisture, drops into a storage bin from which a table feeder discharges to a conveyor system for blending flotation concentrate with the Wetherill concentrate for shipment.

As conditioning pH is critical, a Beckman pH meter and a Bristol recorder are used. The electrode is located at the discharge of the third conditioner. Automatic pH control is anticipated. Feed and tailings are sampled automatically; concentrates and intermediate products are hand sampled. A cut of the tailings is tabled for pilot purposes. Besides assays of shift and composite samples, daily standard flotation tests on the plant-conditioner discharge and on laboratory-conditioned pulp are made. Shift operations are under direction of a technically trained supervisor.

Screen analyses of flotation and regular plant intermediate and final products are given in Table 10.

The metallurgical balance of the flotation operation, based on feed to the conditioners, is given in Table 11. At present, there is some loss of recoverable ilmenite in the overflow of the Dorr rake classifier preceding the conditioners, as the only classifier available was a 5-ft wide unit. Additional settling area, to be provided, will increase the tonnage of flotation feed to approximately 350 tons per day.

TABLE 10—*Screen Analyses of Intermediate and Final Products*

Product	Percentage Weight Retained, Direct							
	+28	+35	+48	+65	+100	+150	+200	−200
Rod mill feed (new)	80.4	3.7	3.3	2.1	2.5	2.3	1.5	4.2
Screen undersize	4.8	11.7	15.4	14.3	13.0	11.7	7.8	21.3
Magnetite concentrate	6.7	15.2	17.6	14.6	11.7	10.3	6.2	17.7
Crockett slimes				1.5	3.5	9.0	12.3	73.7
Dorrco sizer slime				1.0	2.2	8.2	13.5	75.1
Slime thickener feed				2.4	1.6	6.2	11.7	78.1
Slime thickener overflow				0.5	0.9	4.5	2.4	91.7
Flotation feed				0.8	2.1	8.1	20.8	68.2
Total tailing	2.5	6.3	9.9	11.6	13.5	13.4	10.1	32.7
Wetherill ilmenite	4.1	13.7	21.1	19.9	17.1	13.5	7.3	3.3
Flotation ilmenite					0.6	3.7	8.9	86.8

TABLE 11—*Flotation Metallurgical Balance*

Product	Tons	Weight, Pct	Assay, TiO_2, Pct	Content, Tons, TiO_2	Distribution, TiO_2, Pct
Flotation concentrate	95	35.2	46.5	44.1	86.0
Flotation tailing	175	64.8	4.1	7.2	14.0
Flotation feed	270	100.0	19.0	51.3	100.0

The oleic acid and fuel oil used impose problems with rubber-covered wearing parts and with filtration. Neoprene is a partial answer for wearing parts. It does not have the abrasion resistance of natural rubber of proper characteristics with oil-free pulps, but the oleic acid and fuel oil causes rapid deterioration of natural rubber. Stainless steel filter cloth, although not adapted for disc-filter use with standard wood sectors, is markedly superior to ordinary canvas which blinds rapidly. The use of special cloth weaves, woven with synthetic filaments, is being investigated.

The present overall mill metallurgical balance is given in Table 2 (Case 3). Of the total increase in TiO_2 and ilmenite production compared to Case 2 (former results), about 45 pct can be attributed to the cobbing operation, producing higher grade mill feed and higher recovery, about 15 pct to improved gravity concentration results, and about 40 pct to extra production from flotation. With sufficient classifier capacity ahead of flotation, recovery will be increased to 57.6 pct, and tailing losses reduced to 15.9 pct. The present loss in the magnetite, amounting to about 26.5 pct of the TiO_2 in the mill feed, is due principally to ilmenite in solid solution in the magnetite. This problem is illustrated in the micrograph shown in Fig 3, of a particle of magnetite concentrate. This micrograph is from a report prepared by the American Cyanamid Co. covering a quantitative examination of magnetite concentrate. This report concludes that 86 pct of the ilmenite in the magnetite is in the solid solution form.

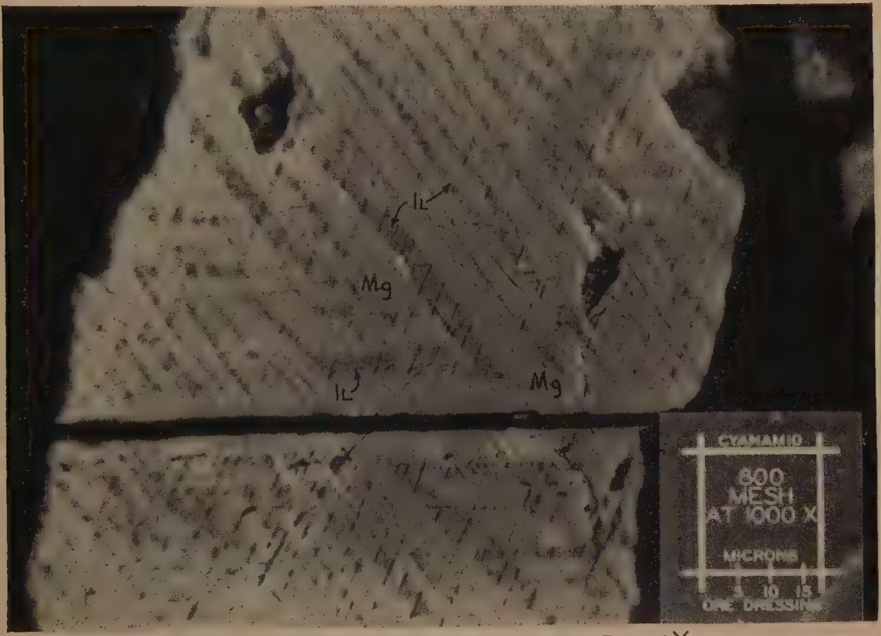

FIG 3—PLUS-200 MESH MAGNETITE PRODUCT. 1000×.

An example of the extremely fine Widmanstatten intergrowth of magnetite and ilmenite is illustrated. Note that many of the narrow streaks of ilmenite are of micron size. The numerous short, straight, black lines are not ilmenite but represent incipient cleavage cracks in the magnetite. (Reduced ⅕.)

Mg—Magnetite (light gray). Il—Ilmenite (dark gray).

Consideration is now being given to the finer grinding flowsheet (to minus-65 mesh) previously referred to. In this flowsheet, Crockett separators would still recover a magnetite product, but the entire Crockett tailing would be treated by flotation for ilmenite recovery. Metallurgically, important improvements in ilmenite recovery and grade and magnetite grade would result. The loss in the magnetite would be reduced from 26.5 pct of the total TiO_2 in the mill feed to about 20 pct of the total TiO_2 (more than is indicated from microscopic examination), leaving 80 pct of the TiO_2 available in the flotation feed. An 85 pct recovery of the TiO_2 available to flotation would result in a 68 pct overall recovery compared with 57 pct now expected with the present flowsheet. This contemplated change would again parallel sulphide metallurgical developments of several decades ago.

ACKNOWLEDGMENTS

During the war years, staff changes at MacIntyre, as elsewhere, were numerous, and many individuals contributed to the various developments discussed. Special mention should be made of C. A. Remington, deceased, formerly metallurgist and mill superintendent, Jay F. Silva, formerly metallurgist, and Professor Charles E. Locke, consultant. The assistance of representatives from various companies, particularly the American Cyanamid Company, the Denver Equipment Company,

and the Dorr Company was also invaluable. The present mill staff, comprising John J. Strohl, Mill Superintendent, Henry J. Schwellenbach, Assistant Mill Superintendent, W. P. Jenkins, Mill Foreman, Robert I. Kingman, Metallurgist, and William Aubrey, Colin Couzens, Thomas MacDonald, Flotation Supervisors, has been responsible for much of the investigative work and reduction to practice.

Gloyd M. Wiles, former Plant Manager, besides participating in the technical work, vigorously prosecuted executive action where required. General company policy of individual responsibility and freedom of action was also an important factor.

DISCUSSION

Nathaniel Arbiter[*]—This paper is an important contribution to milling technology. The findings with respect to the relation between energy input in conditioning and flotation results are particularly interesting.

The statement is too frequently made that any result obtained in laboratory flotation can be duplicated in the mill. If the amendment is added: *Provided that the hydrodynamic conditions can be duplicated,* the statement would be more accurate and less conducive to gray hairs. Because we usually bury our failures in unmarked graves, the literature does not often record the unsuccessful attempts to achieve such duplication.

In connection with one investigation, the outcome of which has not yet been as fortunate as Mr. Milliken's, the following differences between laboratory and mill cells were determined:

Cell	Aeration Intensity Cu Ft/Sq Ft/Min	Ratio, Volume/Area	Air Released Through Froth, Cu Ft/Ton of Ore
Laboratory 2 liter.....	1.61	0.22	3,970
Mill 70 cu ft nom......	2.58	1.79	860

Both cells were of the same design.

It may be true that for simple ores and simple separations such differences are not significant. But where refractory ores are involved, where small but important metallurgical improvements are demanded, or particularly where oxidation in pulp or froth is important, such differences can make laboratory results unrealizable in a mill circuit.

As ores become more difficult, we may have to insist that cell manufacturers adhere to principles of dynamic similarity in designing laboratory and mill cells. Until they do it is well to re-emphasize that the flotation cell or conditioner can be a variable at least as important as reagent concentration, and that laboratory and mill machines do not run by the same clock unless aeration intensity, unit air supply, and turbulence, among other things, are comparable.

C. E. Locke[*]—This paper is of peculiar interest to me because as stated therein I had been connected with various and sundry tests on MacIntyre ore extending over a quarter of a century. However, all tests were on one type of high grade ore and the mill design was based on these tests. It was war time and the need was most urgent so that the mill had to be built immediately to get production. It was impossible to take time for thorough study of the various types of ore and their amenability to concentration and especially to look into the application of flotation which had not been given any previous attention but which seemed to offer much promise.

The long series of tests that had been made clearly indicated a flowsheet combining magnetic separation with gravity concentration on tables and the tests gave the necessary data for developing the details of the mill design. So the mill was built and most happily through the fine work of the designers and the construction people it functioned right from the beginning. A satisfactory ilmenite concentrate although somewhat reduced in tonnage was produced in spite of the changed mill feed. The company had the further good fortune to be able to make the changes outlined by Mr. Milliken without seriously affecting the mill operation. Thus it may be said that the enterprise has come through successfully under conditions where luck seemed to play a role and they had mostly good luck.

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This paper stresses again the old principle that no mill should be built (1) until the mine has been sampled in every part to determine what will be the mill feed, and (2) until all tests have been completed to give the correct mill design. So important is this that it has been the practice of one large company to sample and test thoroughly and then repeat the procedure to check their first results.

Tailings and Mine-dump Reclamation in the Coeur d'Alenes during World War II

By W. L. Zeigler,* Member AIME

(New York Meeting, March 1947)

During the middle 1880's, shortly after the discovery of silver-lead ores in the Coeur d'Alene district of northern Idaho, it became apparent that concentration of the ores would be necessary to obtain a profitable product. Too much siliceous material was present for direct smelting, and the majority of the ores found at depth were too low in metallic content for economic smelting processes at the time. As the ores occurred in quartzite shear zones with siderite and quartz gangue, a large quantity of waste material was present in all the mines. As the galena was not finely disseminated, this made possible relatively good extraction of the argentiferous galena by coarse concentration.

The first concentrators consisted mainly of coarse crushing, screening and jigging, principally by Harz jigs, the slimes being treated on buddles. Later Wilfley tables took the place of sand jigs and vanners superseded the buddles. Because of the marketing conditions, no attempt was made until 1900 to effect a recovery on the zinc minerals present, which were found associated in varying amounts with the galena.

All through this period, the greater part of the production was obtained from mines containing ore assaying high in lead and low in zinc. Parts of some ore bodies would be reversed in this respect, so that ores high in zinc were left intact. During the decade after 1910, several important mines, The Success and The Interstate Callahan, containing ores high in zinc, were put into operation and were a material aid in the needed zinc production during World War I. Gravity concentration by jigs and tables was used during this period, with bulk flotation increasing in popularity for the treatment of the fine material that had already been treated by ordinary slime concentrating machinery. No extra fine grinding for flotation concentration was thought profitable at the time.

Tailings Discharged into Streams

From the start of mining and milling operations in this district until 1904, no attempt was made by the various mills in the upper district to impound or stack their tailings. They discharged them directly into the adjacent creeks or streams, to be carried away by high water. The Bunker Hill and Sullivan and the Last Chance mines, both in the relatively flat valley of the district, stacked huge piles of the coarse jig tailings on the near-by river bottom.

Rain and snowfall are heavy in the Coeur d'Alene district, as it is on the heavily timbered Pacific slope of the Bitter Root Mountains. A heavy runoff occurs every spring and there is an occasional flood of disastrous proportions when warm "Chinook" winds accompanied by rain

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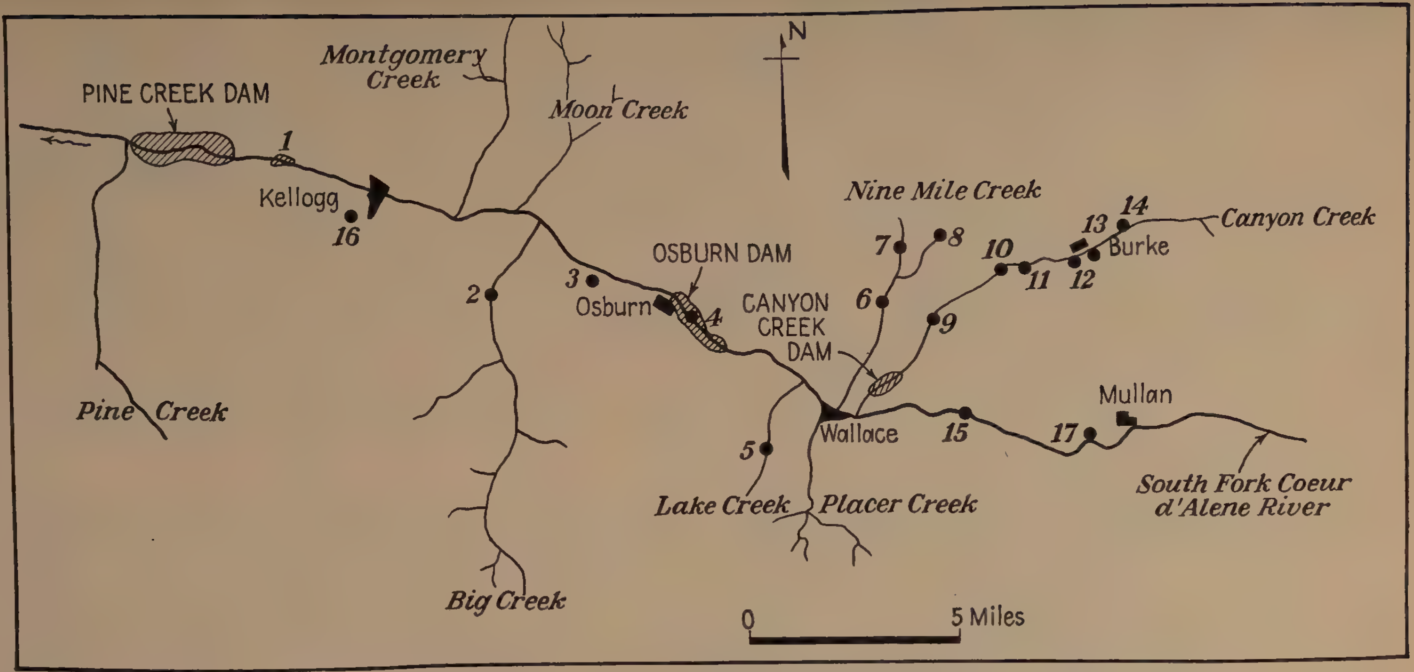

Fig 1—Locations of reclamation projects in Coeur d'Alene mining district.

1. Sweeney mill site.
2. Sunshine mine and mill.
3. Polaris mill.
4. Osburn tailings plant.
5. Galena mill.
6. Success mine.
7. Interstate Callahan mine.
8. Tamarack and Custer mine.
9. Formosa mill.
10. Hecla Gem plant.
11. Frisco mine.
12. Star mill.
13. Hecla mine.
14. Hercules mine and Tiger mine.
15. Golconda mine.
16. Bunker Hill and Sullivan mine and mill.
17. Morning mine and mill.

rapidly melt deep snows in the middle of the winter. Because of this, the coarse mill tailings that were run into the smaller streams during the low-water seasons were carried downstream and spread over areas of low ground on the river bottom.

Dumps Formed

In 1903, litigation was started by farmers in the lower Coeur d' Alene River Valley to prevent pollution of the stream by the mining and milling operations above. This action led to the construction of pile and plank dams across the river beds at three different points in the district; mainly at Canyon Creek, Osburn, and Pine Creek, as shown by Fig 1. After the dams were in place, tailings were held back at the various points, until the flood of 1917, which caused a partial failure in the face of the Canyon Creek and Osburn dams. As the areas behind these dams were already filled with tailings and river rubble, no attempt was made to restore them at the time. Several attempts, on a small scale, were made to rework the material that had been stored behind these dams, but they were only modestly successful and then only during high metal prices. Later fine grinding and flotation concentration was adopted in the mills and the dams were of no further use.

In order to prevent pollution of the lower Coeur d'Alene River Valley, a joint dredging operation was adopted in 1932, on the river near Cataldo, which is at slack water about 15 miles below the mining district. The dredge excavates fine tailings settled from the stream each year, and pumps them for storage on near-by swamp ground where the material is safe from river currents, thus preventing them from being carried on to Coeur d'Alene Lake, about 23 miles below.

Reworking Projects

In 1939, after the beginning of World War II, increasing prices for metal led to consideration of the reworking of tailings deposits as well as mine dumps containing small amounts of minerals. This was made more attractive when the United States entered the war and the quota system of metal subsidies was instituted by the Metals Reserve.

During 1940, skilled tradesmen began leaving the district in favor of the higher wages being paid for war work outside of the Coeur d'Alenes. High wages were paid for all types of labor on the construction of the Farragut Naval Station, which is less than 100 miles distant, and the mines suffered severely from the loss of men to this work. About this time, Velox Naval Station, near Spokane, was started, and was followed immediately by the construction of the Alcoa aluminum plant and rolling mill, as well as the magnesium plant, Geiger Air Field, Galena Air Base, and lastly, the secret Hanford atomic bomb plant near Pasco. Underground production of ore at the mines dropped to about one third of normal tonnage, and in some cases considerably more.

Bunker Hill and Sullivan Mills

As tonnage of mine ore dropped at the Bunker Hill and Sullivan mine at Kellogg coarse jig tailings that had been stacked at the old Sweeney mill site were hauled to the West mill of the Bunker Hill,[1] and treated as a mixture with the mine ore. A sink-float plant had been installed in 1941 to beneficiate the ore previous to selective flotation concentration, and it proved equally beneficial for the treatment of the tailings. At first about 500 tons per day of these jig tailings was excavated for milling but the amount was gradually increased until more than 1000 tons daily was being milled.

After the Sweeney tailings were exhausted, operations were shifted to the vast pile of jig tailings left from the

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[1] References are at the end of the paper.

Bunker Hill's West mill operation dating from 1908. Also, a concentrator[2] of 300 tons daily capacity was built in 1941 to produce a high-iron flux for the lead smelter. Jig tailings stored in a separate pile from some of the original milling are excavated and crushed to ½ in., screened into three sizes, and concentrated by Harz jigs and tables. Resulting concentrates have an iron content above 28 pct and contain about 50 pct of the lead and silver in the original tailings. A coarse concentrate is imperative for flux because of the needed porosity in roasting. The tailings from this concentration are then ground and treated by flotation for a silver-lead concentrate only. This plant operates eight months of the year, which meets the yearly flux demands of the smelter.

Interstate Callahan and Galena Mills

In 1940, Zanetti Bros. constructed a coarse-jigging plant at the site of the razed Interstate Callahan mill. Some remaining jig tailings were re-treated and a large dump of hand-sorted waste containing 0.4 oz Ag, 1.0 pct Pb and 2.0 pct Zn was worked, along with a considerable tonnage from the main mine dump. The combined amount of material treated was about 100,000 tons. The rough jig concentrate was hauled by truck to the Galena mill,[3] in Lake Gulch below Wallace. This mill has a capacity of about 150 tons daily and is equipped to produce both lead and zinc concentrates by flotation. Zanetti Bros. also worked several areas along the south fork of the Coeur d'Alene River, between Wallace and Mullan, with a portable screening plant, to eliminate river rock and rubble.

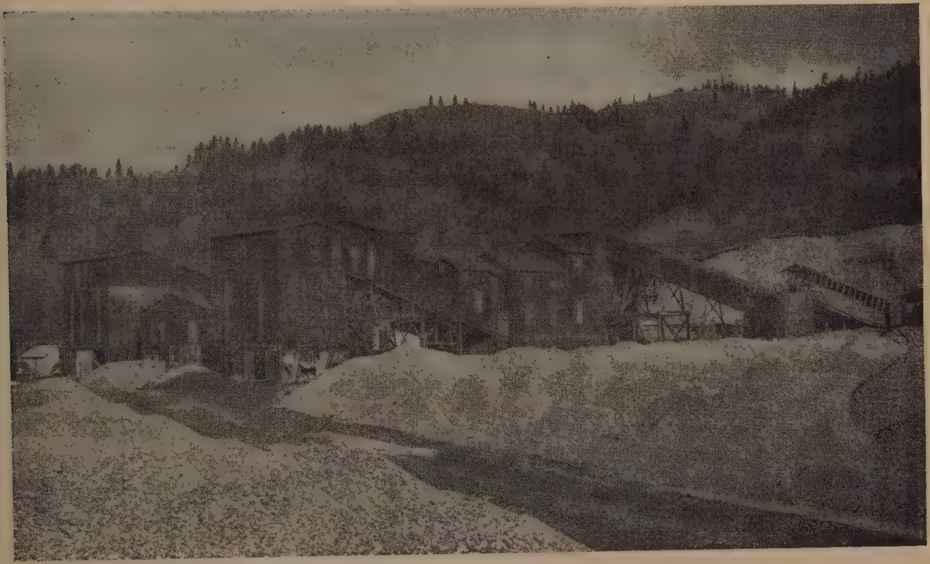

FIG 2—OSBURN TAILINGS PLANT, HECLA MINING COMPANY.
Float or waste piled in foreground; screened rubble at right center.

Another interesting project of the Zanettis was the beneficiation of 400,000 tons of hand-sorted waste that had been stored for many years at the upper portal of the Tamarack and Custer mine. This waste, assaying 0.4 oz Ag, 1.5 pct Pb and 0.5 pct Zn, was picked from sorting belts when the ore was milled at several old plants. The ore from the upper workings contained a variable quantity of magnetite, which interfered with jigging operations, as it lowered the grade of the lead concentrate. In the reworking of the dump, the material was excavated and simply passed over a magnetic pulley, the nonmagnetic portion going to waste. The magnetic material contained over 50 pct of the lead and silver and assayed roughly 5.0 pct Pb and 1.5 pct Zn. This was then hauled to the Galena mill for conventional treatment.

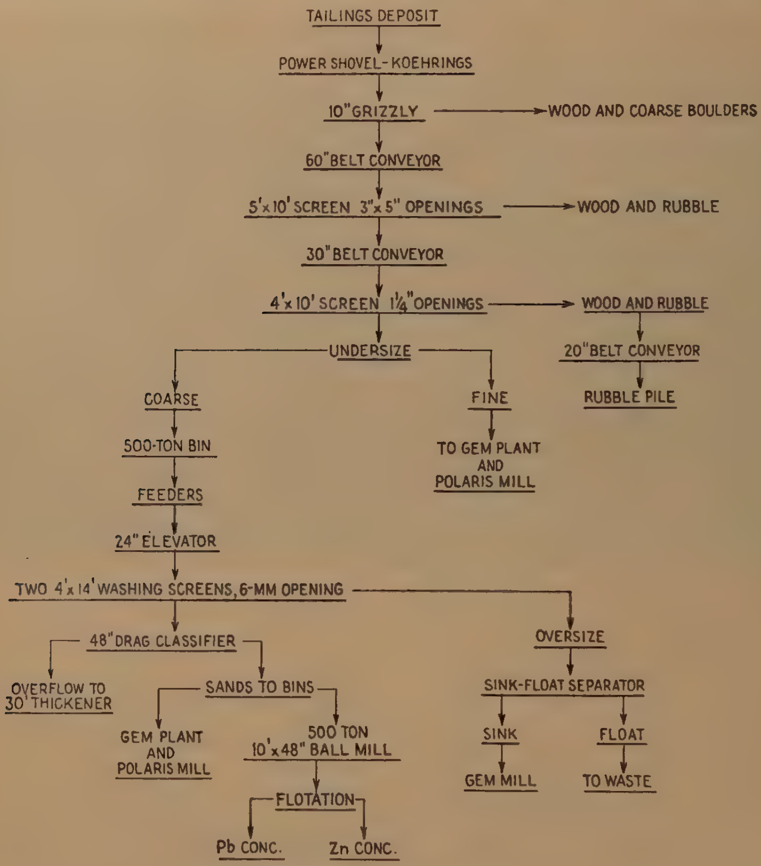

Fig 3—Simplified Osburn flowsheet.

Morning Mill

At the portal of the main tunnel of the Morning mine at Mullan, a very large waste dump had accumulated. Because of the great depth of the Morning workings, this mine suffered a greater labor shortage proportionally than any other in the district, and the tonnage from underground operations dropped rapidly. This was made up in part by material excavated from the dump by power shovels and trucks. About 500,000 tons, assaying 0.4 oz Ag, 1.1 pct Pb and 0.6 pct Zn, was treated in the Morning mill[4] along with the regular mine ore.

Golconda Customs Mill

The tailings impounded above the Canyon Creek dam, about one mile above Wallace, were higher in zinc content than the average of the accumulations at other points, because when the Tiger and Frisco mills above were operated a lead-zinc ore was treated but a marketable zinc concentrate could not be produced by methods then in use. This deposit, about 200,000 tons, was worked for a short time in 1937 on a small scale, and was treated at the Golconda customs mill, which has a capacity of 175 tons per day.

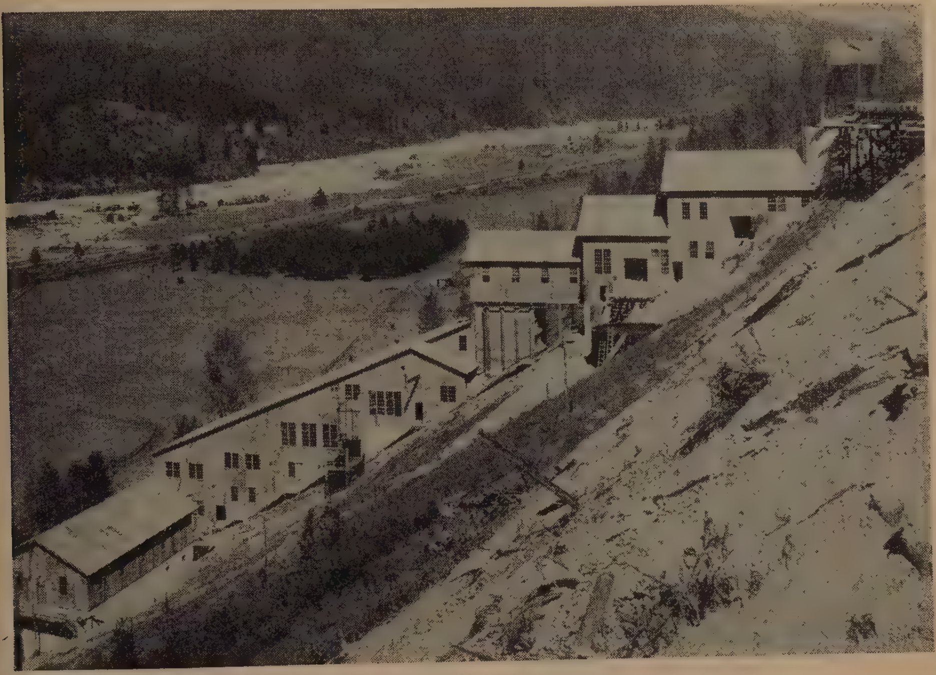

FIG 4—POLARIS CONCENTRATOR.
Looking up the valley of the south fork of Coeur d'Alene River. Tailings are shown along the River at foothills across the valley. Osburn dam is up the valley at the extreme right of the picture.

Formosa Mill

In 1942, the Small Leasing Co.[5] reinstalled the machinery in the old Formosa mill, about one mile above the deposit. It was equipped to produce both a lead and a zinc concentrate, with an initial capacity of 125 tons per day. After several years operation on the higher grade tailings, a jigging plant consisting of screens and a Bendelari jig was installed at the site of the deposit to beneficiate tailings from clean-up areas that were too low grade for the regular treatment methods. This plant did not prove economical and a sink-float plant of 400 tons daily capacity was later installed at the mill. Float, or the waste portions of the tailings, is discarded, while the sink and sands are treated at capacity in the grinding and flotation circuit of the mill.

Pine Creek Dam Area

Several attempts have been made in the past to rework tailings settled from the river behind the Pine Creek dam. As both the Bunker Hill and Sullivan and the Sweeney mills stacked their jig tailings near this site, and the Osburn dam held back most of the coarse tailings from the mills above, the material in this area is very fine. Because of this, the lead minerals present are almost completely oxidized, which complicates selective flotation and makes it difficult to produce a marketable concentrate of either lead or zinc, as well as economical recovery.

In 1941, the Gibbonsville Mining Co. erected a plant of 250 tons daily capacity to treat the fine material from this area. Very little grinding was necessary, so the mill consisted mainly of conditioning and flotation equipment. A satisfactory recovery of the metals was not obtained, and after several runs were made by different operators, the plant was finally dismantled.

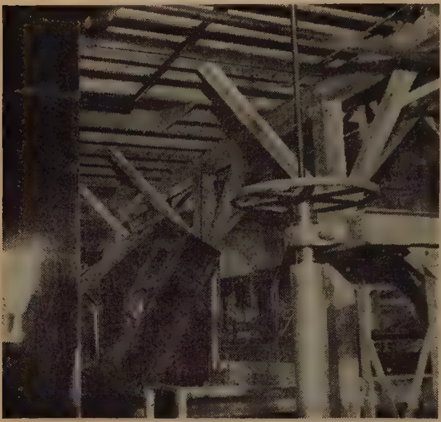

FIG 5—ELEVATORS AT OSBURN.
Medium elevator at left; top of sink elevator at center. Medium storage tank at right.

Hecla Mining Company Mills

Early in April 1943, the Hecla Mining Co. placed in operation on a small scale a plant at Osburn, Idaho,[6] designed to beneficiate eventually 3000 tons daily of a large accumulation of jig tailings, conservatively estimated at 3,000,000 tons, which had been impounded by the Osburn dam (Fig 2). The tailings were from the Gold Hunter, Morning, Snowstorm, National, Hercules, Frisco, Hecla, Standard, Mammoth, Interstate, Success, Tamarack, and Tiger mines, the respective mills being situated on the streams above the dams. This plant, which has proved to be by far the most important reclamation project in the district, now produces a feed of commercial grade for the Company's different flotation mills, as well as lead and zinc concentrates for smelting. The plant methods employed include screening, sink-float beneficiation, and lead and zinc flotation.

The material treated at the plant is mined by power shovels and transported to the plant by 12-yd Koehrings, which are bottom-dump semitrailers powered by two-cycle G.M. diesel engines (Fig 3). Bulldozers assist the shovels by stockpiling ahead of them. The treatment plant proper is at about the center of the deposit, which is approximately one mile long and one half mile wide, and which is readily accessible from the main highway, about 500 ft distant. The material from the tailings deposit is dumped into a 30-ton steel-lined loading bin at the plant, the top being accessible by two earth ramps. A steel grizzly on top of this bin, having 10-in. openings, separates the coarse river gravel and pieces of driftwood from the tailings, and these are then removed by hand after each load is dumped. The undersize, or tailings, amounting to 3000 tons per day, travels up a 60-in. belt conveyor to a 5 by 10-ft Ty-Rock screen having 3 by 5-in. openings, where more wood and coarse rock are removed as oversize and transported to a bin by a 42-in. belt conveyor. This waste material is hauled by truck to the worked-over areas of the tailings deposit.

The undersize from the 5 by 10-ft Ty-Rock travels up an inclined 30-in. belt conveyor to the top of the screening plant. Here it is screened on a 4 by 10-ft Ripl-flo screen having 1¼-in. openings from which the undersize falls directly into the 500-ton main storage bin. The oversize is carried away by a 20-in. belt conveyor and stacked on the rubble pile at the rear of the plant. Fine material falling into the bin from the feed end of this screen, being higher in metallic content than the average tailings, is removed by a side chute and hauled to the Gem and Polaris (Fig 4) mills without further processing. Trucks are used entirely for hauling to the Polaris mill, which is about 1½ miles west of the Osburn plant. The material for the Gem mill, which is approximately 9 miles from the tailings deposit, is loaded by trucks at the Osburn siding into 80-ton air-bottom-dump railroad cars. At Gem it is dumped into the ore bins.

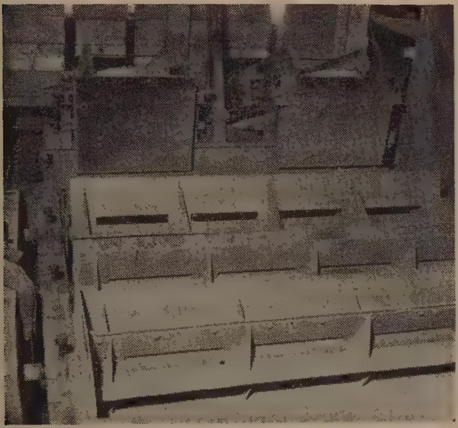
FIG 6—TOP OF 9 BY 8-FOOT SINK-FLOAT SEPARATOR.
Storage bin and feed conveyor above

The coarse material accumulates to a large degree on the other side of the bin and forms the true feed to the plant. From here it is withdrawn and elevated to two 4 by 14-ft Lowhead washing screens having 6-mm openings. The oversize produced here (−1¼-in. + 6 mm) forms the feed to the sink-float separator (Figs 6 and 7), which is of the Huntington-Heberlein type, using galena as medium. The float product is washed and conveyed to a bin, from which it is hauled away and discarded. The sink product undergoes the same treatment and is hauled to the Gem mill together with the other products.

The undersize from the 6-mm screens goes directly to a 48-in. drag classifier, where a separation is made at about 65-mesh. The minus 65-mesh material constitutes the overflow and discharges by gravity into a 30-ft thickener. The rake product (−6-mm + 65-mesh) is transported by a belt conveyor to a split bin; one section of which is used as a feed bin for the ball mill while the other is used as a loading bin, from which the sands are hauled to the Gem and Polaris mills by the methods described earlier.

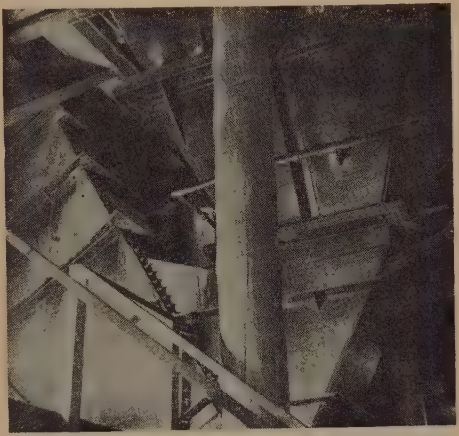
FIG 7—BOTTOM OF SINK-FLOAT SEPARATOR.
Sink elevator housing at left.

The ball-mill circuit consists of a 10-ft by 48-in. Hardinge ball mill in closed circuit with a Wemco 50-in. duplex submerged spiral classifier (Fig 8). The classifier overflow is pumped to an eight-cell bank of 56-in. Fagergren flotation machines, of which four are used as lead roughers and four as zinc roughers. Both rougher concentrates are cleaned twice in 18-in Sub-A flotation machines. The concentrates are filtered and stored for shipment to their respective smelters at Kellogg.

In planning this project, factors such as the limited life of the operation, scarcity of appropriate building materials, and lack of skilled and unskilled labor, had to be taken into consideration. Inability to obtain steel prevented its use in the construction of the plant buildings; and the Navy order freezing cement and sawed lumber came during the initial planning stages, which led to the adoption of wood-pile foundations for all of the construction; 347 pilings were driven to a depth of 6 to 8 ft in the river gravel by an air-driven pile hammer, for the various loading and storage bins, thickeners, and the first story of the buildings proper. Because of the scarcity of timber, ordinary round mine stulls were used for this work. All the buildings and bins are wood, with roofs and exterior walls covered with asphalt composition roofing and felt paper, respectively.

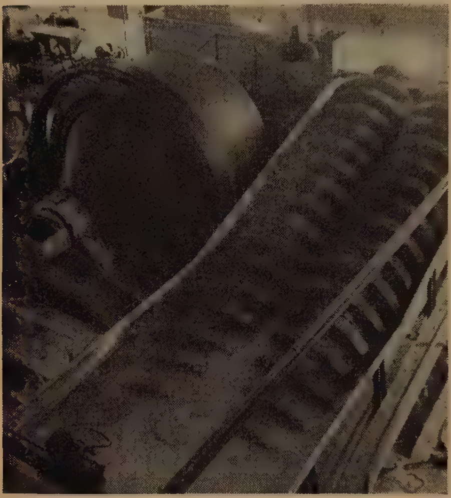

FIG 8—GRINDING UNIT, OSBURN TAILINGS PLANT.

Hardinge ball mill and submerged classifier. Dry reagent feeders at upper left.

The production of silver, lead, and zinc from the Osburn plant in the year 1945 ranked high with that of various mines in the state of Idaho. It was exceeded in production of silver only by the Sunshine mine, which is a silver mine; in lead only by the Bunker Hill and Sullivan mine, which is essentially a lead mine; and in zinc only by the Star mine, the ore of which contains mainly zinc. Because of the continuation of the shortage of underground labor in the district, the same will probably be true of the production of metals in 1946.

Recovery of Metals

In most of the old gravity-concentration plants in the district, ores were crushed to pass 30-mm round-hole screens before jigging was started. Consequently, in all the recent tailings reclamation projects, foreign material over this size, such as river gravel and wood, are screened out as waste before treatment is begun. Oxidation of the galena is apparent in all the old tailing deposits in varying degrees, being only superficial on coarse sizes and increasing to complete oxidation of sizes finer than 65-mesh. For this reason, recovery of contained lead and silver is low, as no economical method of flotation concentration has been devised to recover the oxidized material. An average recovery of the metals has been approximately 65 pct of the silver, 50 pct of the lead, and 75 pct of the zinc (Table 1).

The only large deposits of material remaining in the district that may be reworked in the future are the huge Bunker Hill and Sullivan West-mill jig-tailings pile, and the remainder of the tailings in the Osburn dam area, which may amount to about 1,000,000 tons.

Ordinarily concentrates produced by treatment of tailings by flotation have a metallic content comparable with those from current higher grade mine ores. The silver-lead concentrates have a typical assay of 45 oz Ag, 60 pct Pb and 7 pct Zn, while zinc concentrates average about 5 oz Ag, 3.5 pct Pb and 50 pct Zn.

Thirty years ago the concentration of ores was considered an individual problem for each mine, and the flowsheets of the mills varied widely, according to the grades and characteristics of the ores. At the present time, a universal method of

TABLE 1—*Summary of Projects*

Operation	Tons Treated	Assays			Metallic Content		
		Ag, Oz	Pb, Pct	Zn, Pct	Ag, Oz	Pb, Tons	Zn, Tons
Hecla, Osburn	2,700,000	1.3	1.7	1.7	3,510,000	45,900	45,900
B. H. and S.: Sweeney[a]	426,000	0.7	1.8	0.8	298,200	7668	3408
West mill	450,000	0.5	1.2	0.6	225,000	5400	2700
Iron plant	280,000	0.9	1.9	0.9	252,000	5320	2520
Zanetti: Interstate[a]	500,000	0.4	1.0	2.0	200,000	5000	10,000
Tamarack[a]	400,000	0.4	1.5	0.4	160,000	6000	1600
Misc. proj.[a]	50,000	0.5	1.8	1.5	25,000	900	750
Morning, mine dump[a]	500,000	0.4	1.1	0.8	200,000	5500	4000
Small Leasing[a]	200,000	0.6	1.9	2.4	120,000	3800	4800
Miscellaneous (Golconda mill)[a]	100,000	1.0	1.5	2.0	100,000	1500	2000
Grand total	5,606,000				5,090,000	86,988	77,678

[a] Exhausted in 1946.

treatment by flotation of the composite tailings of all the old mills is used, without any appreciable difficulty.

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The Flotation of Fluorite

By Enid C. Plante*

(New York Meeting, March 1947)

This paper deals with the flotation of the mineral fluorite (calcium fluoride) and of two associated gangue minerals, calcite and quartz. The aim of the investigation was to produce "acid-grade" fluorite, which must contain less than one per cent of silica.

Several reagents are used successfully in practice for floating fluorite from its ores. Oleic acid is the most widely used collector, usually at temperatures above 60°C, with sodium silicate as a dispersant for quartz and quebracho or tannin to depress calcite. Sulphides (e.g., galena) are often removed first by xanthate flotation. Many investigators[23] report results in which more than 80 pct of the fluorite is recovered in a concentrate assaying 96 pct or more of fluorite and less than 1 pct of silica.

This investigation was undertaken with the intention of floating the gangue, which is the minor constituent of the ore, from the fluorite. Hence emphasis was first placed on the cationic compounds that float quartz. No satisfactory separations were obtained in this way, and other reagents were investigated to discover whether these were more effective than oleic acid in floating fluorite from quartz in the usual way.

The first part of the paper describes experiments with several long-chain collectors available commercially. Most of this work has been by captive-bubble and cylinder flotation tests, with some batch tests in a flotation machine. The second part describes the use of sodium cetyl sulphate as a collector for fluorite and includes laboratory tests and extension of the work to batch tests in the flotation machine. A method for determining the adsorption of sodium cetyl sulphate on fluorite has been developed and is described.

Theory of Collection by Paraffin-chain Salts

It may be assumed that these compounds are adsorbed at the mineral surface in one or more of three ways outlined by Kolthoff:[12]

1. Exchange adsorption between lattice ions in the mineral surface and collector ions in solution. Closely related is adsorption by the collecting mechanism postulated by Taggart,[17] in which there is metathesis between the collector and the mineral cation or anion.

2. Exchange adsorption between adsorbed gegen ions (p. 67, ref. 19) and collector ions in solution.

3. Adsorption of nonelectrolytes (polar-nonpolar) without displacement of adsorbed ions.

With cationic collectors, inorganic cations in the solution (e.g., hydrogen ion) compete with the collector ions and therefore act as depressants, while inorganic anions have the same effect with anionic collectors. From this follows the general conclusion that ions of the same charge as

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the collector usually act as depressants for that collector, while ions of the opposite charge may be activators. These effects may be partly or wholly prevented by precipitation of the added ions by the collector.

Effect of Collector Concentration on Flotation

Curves have been constructed from captive-bubble and cylinder flotation tests showing the solution compositions in which adsorption of collector and flotation of the mineral occur. The solution compositions in which flotation occurs are distinguished by a full line from those in which it does not, and solutions in which there is adsorption of collector but no flotation are distinguished by a broken line (Figs 1–13). The conditions under which flotation, adsorption and nonflotation occur are marked on figures in which there is only one graph. It would be confusing to show these points on figures containing more than one graph. Cessation of flotation in solutions of high ion concentrations despite the presence of an adsorbed collector film on the mineral is characteristic of most polar-nonpolar compounds[15] and the longer the chain length of the alkyl compounds, the lower the concentration of collector at which this effect occurs (cf. laurylamine and cetylamine, Figs 7 and 8). The phenomenon is postulated to be due to two effects:

1. "Armouring" of the bubble by collector ions.[20]

2. Formation of a partial or complete second layer of adsorbed collector molecules on the mineral, oriented the wrong way for attraction between the collector film and an air bubble (Held and Samakhvalov.[7] Gaudin and Rizo-Patrón,[5] Taggart and Arbiter[16,17] Rogers et al.[15]

This cessation of flotation is induced by sufficiently high concentrations of the collector itself, and it is favored by the presence of neutral salts, acids and alkalies in proportion to their concentrations. Thus the range of flotation is decreased by addition of alkalies and acids apart from any specific effects due to hydrogen or hydroxyl ions.

It should be realized that results obtained in the laboratory by captive-bubble and cylinder flotation tests are not necessarily identical with those obtained in the flotation machine. Many factors other than those considered in these tests influence recovery and grade in practice. Hergt et al.[8] have shown in the flotation of cassiterite that discrepancies exist between laboratory and flotation-machine tests. However, by avoiding marginal (i.e., conditions shown in the figures to be close to the curve separating the region of flotation from that of nonflotation) conditions for separations based on laboratory results, serious discrepancies usually do not occur.

Wark and Wark[21] have shown that discrepancies sometimes exist between captive-bubble and cylinder tests for the response of sulphide minerals to amines. No differences were found for those results reported in this paper, which were checked by both methods. However the discrepancies if they occur are usually small and should not concern the operator who avoids marginal conditions when translating laboratory results into practice.

Experimental Procedure

Preparation of Mineral

For captive-bubble tests mineral specimens as pure as possible were prepared in the manner described by Wark.[19]

In both captive-bubble and cylinder tests the mineral was conditioned for 30 min. in the test solution at 35°C, and then tested for air-bubble contact or flotation. If there was no response by the mineral it was placed in water of the same pH value as the original solution and tested for the presence of an adsorbed film of collector.

For cylinder flotation tests,[19] the mineral was boiled with sodium hydroxide to remove grease, with acid to remove alkali, and finally with water; it was then ground under water with a clean pestle and mortar, washed into a clean stoppered 50-ml cylinder, and shaken with water containing a frother (terpineol) only. Only if no mineral floated was it considered uncontaminated and suitable for tests.

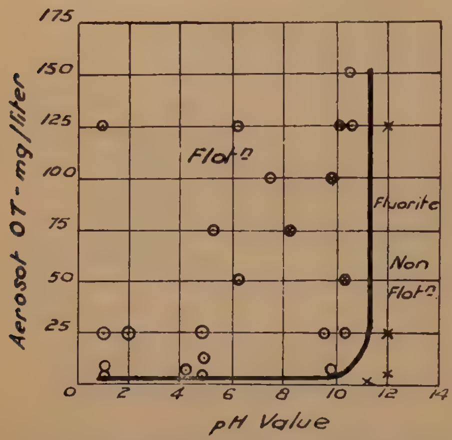

Fig 1—Relationship between Aerosol OT concentration and pH value—fluorite. Temperature, 35°C.

In this and following figures:

Circle indicates air-bubble contact or flotation.

x in circle indicates doubtful contact or flotation.

X indicates no contact or flotation.

Preparation of Test Solutions and Measurement of their pH Value

The test solutions were prepared by adding reagents in the order: (1) acid or alkali to give the desired pH value; (2) activator or depressant if required; (3) collector. The solution was then made to an accurate volume and, for reasons cited by Rogers et al.[15] the pH value was measured with the glass electrode. The pH value of the solution was again measured after the 30-min. conditioning period.

Analysis of Flotation Products

The flotation products were sampled by riffling, the analytical sample ground to minus 100-mesh (to hasten solution) and analyzed by the method given by Low et al.[22] Calcite was estimated, although it is not recorded in the tables and was found chiefly in the tailings. Various amounts of iron were found in the tailings also.

COLLECTORS INVESTIGATED

The following collectors were used: Aerosols MA and OT, Sapamine MS, Ninol 737, Igepon T, cetyl trimethyl ammonium bromide, and laurylamine and cetylamine hydrochlorides. The first five reagents were commercial samples and were used without purification. Cetyl trimethyl ammonium bromide was prepared as described previously[15] and laurylamine and cetylamine hydrochlorides were prepared and purified in the laboratory.

The *Aerosols,* being sulphonates, are adsorbed on the cations of the fluorite surface—i.e. on the calcium ions—and compete with other anions in solution. *Aerosol OT* is the sodium salt of the dioctyl (2-ethyl hexyl) ester of sulphosuccinic acid:—$Na_2O_3S.CHOOC_8H_{17}.CH_2$-$COOC_8H_{17}$.[13] In *Aerosol MA* hexyl (methyl amyl) groups replace the octyl groups of Aerosol OT. McBain and Merrill[13] state that the Aerosols, which are manufactured by the American Cyanamid and Chemical Corp., are 99 pct pure.

There is no significant difference between the curves in Figs 1 and 2, showing the solution compositions in which fluorite is floated by the two compounds. Neither reagent floats calcite or quartz. No anionic compounds are known to be collectors for quartz unless it is activated by metallic ions such as Pb^{++} or $Fe^{+}{}_{+}{}^{+}$.

Surface-tension measurements under similar but not necessarily equilibrium conditions on solutions containing 50 mg of Aerosol OT and Aerosol MA per liter gave values of 44.0 and 56.9 dynes per centimeter, respectively.

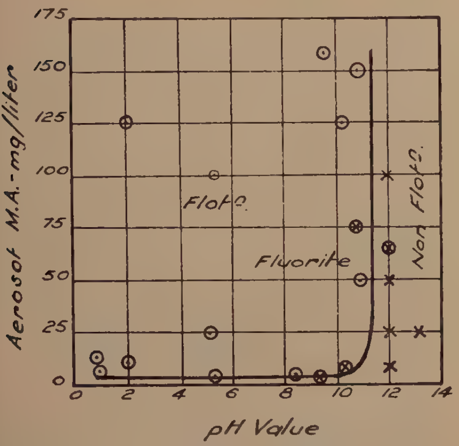

FIG 2—RELATIONSHIP BETWEEN AEROSOL MA CONCENTRATION AND pH VALUE—FLUORITE. TEMPERATURE, 35°C.

It is of considerable interest that repetition of the octyl and hexyl groups in the Aerosols confers on them good collecting properties. Similar compounds with a single ester group would not be expected to be collectors at concentrations as low as 10 mg per liter. Thus Hansen (Table 3, ref. 6) has shown that 1–18 lb of heptylic acid ($C_6H_{13}COOH$) or 0.8 to 4 lb of pelargonic acid ($C_8H_{17}COOH$) per ton of mineral is required to float various carbonate minerals, which is at least 10 times the amount of Aerosol OT or Aerosol MA required to float fluorite.

Igepon T is an anionic collector manufactured by the General Dyestuff Corp. and Colgate Palmolive Peet Co., and has the formula $C_{17}H_{33}CONHC_2H_4SO_3Na$.[2] It is a collector for fluorite, being adsorbed over a pH range from 1 to 13. At high pH values the inorganic ion concentration hinders flotation despite the presence of an adsorbed collector film (Fig 3). Igepon T in concentrations greater than 30 mg per liter floats calcite, there being no detectable difference between flotation and adsorption curves. Igepon T is not a collector for quartz.

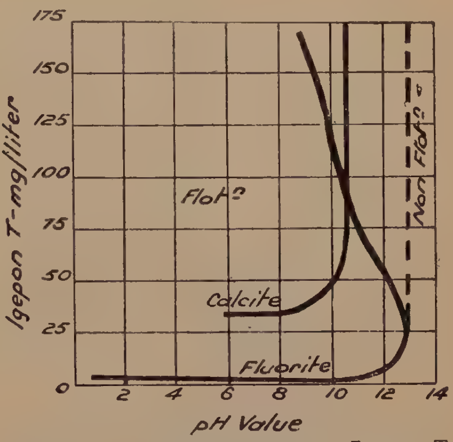

FIG 3—RELATIONSHIP BETWEEN IGEPON T CONCENTRATION AND pH VALUE—FLUORITE, CALCITE. TEMPERATURE, 35°C.

Ninol 737, prepared by the Ninol Laboratories, has been described[10] as a C_{10}–C_{16} alkylolamine condensate and is therefore a cationic collector. In stating that Ninols are collectors only for activated minerals, Dean and Ambrose (p. 31, ref. 4) presumably were referring to anionic Ninols, for later (p. 65) they show that acylated and alkylated Ninols, which are referred to as DLT reagents, are cationic collectors. Fig 4 shows that Ninol 737 is a collector for calcite and quartz up to pH values of about 8 and for fluorite up to pH 13. The nonflotation of calcite and quartz above pH 8 must be due to hydroxyl ions, for both minerals floated in a neutral solution of N/1000 sodium chloride, indicating that sodium ion is not the depressant. Flotation tests in a 500-gram machine confirmed the behavior of calcite. With the addition of 30 mg of the collector per liter, the recovery was 72 pct at a pH value of 7; 44 pct at pH 8 and 4 pct at pH 9. As the reagent is cationic and therefore adsorbed on the anions of the mineral lattice, similar behavior of calcite and fluorite would not be expected, and Fig 4 shows that the regions of flotation differ.

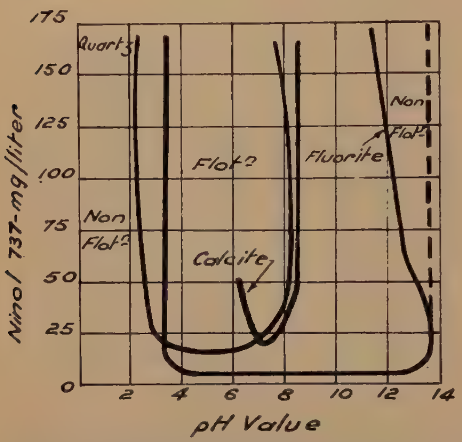

FIG 4—RELATIONSHIP BETWEEN NINOL 737 CONCENTRATION AND pH VALUE—FLUORITE, CALCITE, QUARTZ. TEMPERATURE, 35°C.

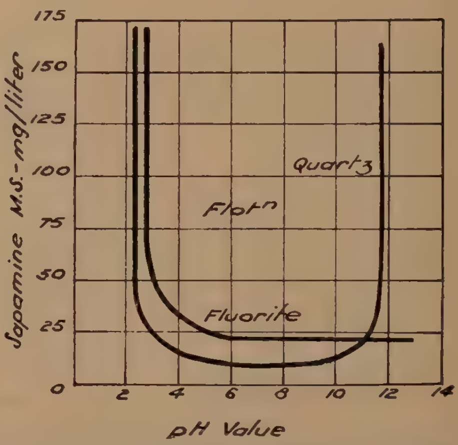

FIG 5—RELATIONSHIP BETWEEN SAPAMINE MS CONCENTRATION AND pH VALUE—FLUORITE, QUARTZ. TEMPERATURE, 35°C.

Sapamine MS is a quaternary ammonium salt-diethyl amino ethyl oleyl amide methyl sulphate.[13] It ionizes in solution and the cation is the collector. It is a collector for fluorite and quartz and for the concentrations tested the conditions for adsorption are the same as those for flotation for each mineral (Fig 5). As much as 175 mg of the collector per liter can be tolerated by fluorite without any diminution of flotation even at a pH value of 13, which is unusual. Sapamine MS does not float calcite.

Cetyl trimethyl ammonium bromide is a cationic collector, being a quaternary ammonium salt [$N.(CH_3)_3.C_{16}H_{33}$]Br. It is completely ionized. Fig 6 shows that it is a collector for fluorite and quartz but not for calcite. Flotation in solution ceases for fluorite when the collector concentration reaches 80 mg per liter and for quartz at 150 mg per liter. This difference might be thought to be due to the higher solubility of fluorite compared with that of the silicates, with a resultant larger amount of electrolyte in solution, an effect that might be partly offset by the greater ionic strength in the quartz pulp of polyvalent silicate ions. However, no confirmation for this view was obtained from a cylinder flotation test with quartz with 100 mg of cetyl trimethyl ammonium bromide per liter, when distilled water was replaced by the filtrate from the pulp of distilled water and ground fluorite which had been kept at 35°C for 30 min.; under these conditions the quartz floated well.

Figs 7 and 8 show the flotation and adsorption curves for *cetylamine* and *laurylamine hydrochlorides*. Both are collectors for all three minerals. With laurylamine the solutions in which adsorption and flotation occur are identical for each mineral, and the solution composition in which collector concentration prevents contact is still not reached at concentrations of collector of 175 mg per liter. Cetylamine hydrochloride, however, with a longer chain length, shows this effect well. Flotation of calcite ceases at 170 mg of collector per liter. With fluorite there is no flotation above 135 mg of collector per liter. The difference between the two minerals can scarcely be due to different solubilities of the minerals, for their solubilities are very similar: calcite = 0.0014 gram per 100 cc at 25°C; fluorite = 0.0017 gram per 100 cc at 26°C.[9] The divalent carbonate ion might be expected to be more effective in solution than the monovalent fluoride; however, two fluoride ions may dissolve for one carbonate, which would tend to counterbalance this effect, or the fluoride and carbonate ions may dissolve at different rates.

The behavior of calcite cannot be studied below pH 6, because solutions with lower pH values dissolve the calcite and the pH value is soon restored to 6 or 6.4.

In solution amines react as follows:

$$RNH_2 + H^+ \rightleftharpoons RNH_3^+$$

so that the relative amounts of free base and ion depend on the pH value of the solution.

The values shown in Table 1 have been calculated by well-established methods, accepting the value of 13.99[11] for *Kw* and calculating *pKa* to be 10.6,* and assuming complete solubility of laurylamine hydrochloride over the whole pH range. However, there is not complete solubility of laurylamine hydrochloride over the whole range of pH values. Taggart, Arbiter

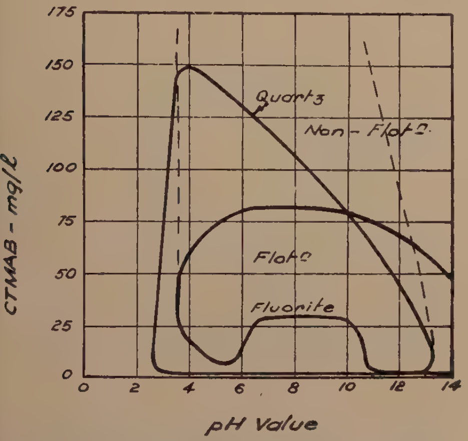

FIG 6—RELATIONSHIP BETWEEN CETYL TRIMETHYL AMMONIUM BROMIDE CONCENTRATION AND pH VALUE—FLUORITE, QUARTZ. TEMPERATURE, 35°C.

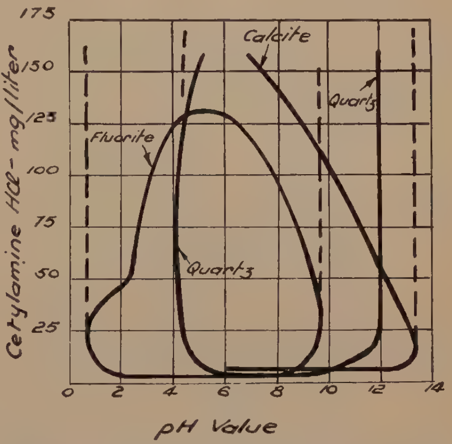

FIG 7—RELATIONSHIP BETWEEN CETYLAMINE HYDROCHLORIDE CONCENTRATION AND pH VALUE—FLUORITE, CALCITE, QUARTZ. TEMPERATURE, 35°C.

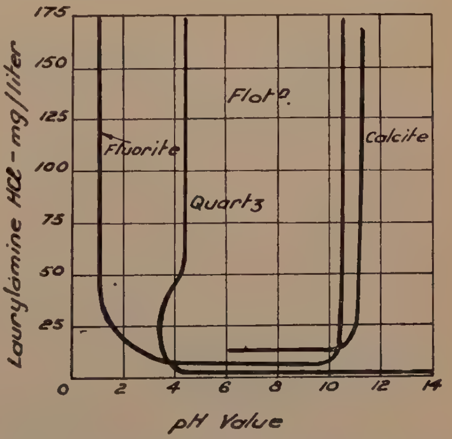

FIG 8—RELATIONSHIP BETWEEN LAURYLAMINE HYDROCHLORIDE CONCENTRATION AND pH VALUE—FLUORITE, CALCITE, QUARTZ. TEMPERATURE, 35°C.

* The *pKa* value of 10.6, based as it is on the accurate conductimetric method used by Ralston, McCorkle and Hoerr[14] is preferred to that given by Taggart, Arbiter and Kellogg;[16] namely 8.8. The former figure is more in line with that for similar amines. In a paper published since this manuscript was prepared, Kellogg and Vasquez-Rosas similarly accept this figure.

and Kellogg[16] report the formation of a Tyndall cone at a pH value of about 8.5, which means that the solubility of free base must be just below 0.2 mg per liter at this pH value. Hence in Table 1 the figures for the ratio of base to ion are valid, and columns 3 and 4 give correct values up to pH 8.5, but above pH 8.5 the values are far too high.

TABLE 1—*Concentration of Laurylamine and Laurylammonion Ion in Solution at Different pH Values*

FOR SOLUTIONS CONTAINING 25 MG LAURYLAMINE HYDROCHLORIDE PER LITER

pH Value	$\frac{C_{12}H_{25}NH_2}{C_{12}H_{25}NH_3^+}$	$C_{12}H_{25}NH_2$, Mg per Liter	$C_{12}H_{25}NH_3^+$, Mg per Liter
1	$\frac{2.38 \times 10^{-10}}{1}$	5.0×10^{-9}	21.0
4	$\frac{2.38 \times 10^{-7}}{1}$	5.0×10^{-6}	21.0
8	$\frac{2.38 \times 10^{-3}}{1}$	5.0×10^{-2}	21.0
8.5	$\frac{2.38 \times 10^{-2}}{\sqrt{10}}$	0.2 (0.16)	20.8
9	$\frac{2.38 \times 10^{-2}}{1}$	0.50	20.5
10	$\frac{2.38 \times 10^{-1}}{1}$	4.0	17.0
10.6	$\frac{1}{1}$	10.5	10.5
11	$\frac{2.38}{1}$	14.8	6.2
12	$\frac{2.38 \times 10}{1}$	20.2	0.8
13	$\frac{2.38 \times 10^2}{1}$	20.9	0.09

Table 2 shows the effect of this restricted solubility on the amounts of free base and ion actually present in solution between pH 8.5 and 13, it being assumed that the base has a solubility of 0.2 mg per liter over the pH range of 8.5 to 13.

The concentration of ion in solution is appreciable up to pH 10.6. Captive-bubble tests with polished quartz specimens at pH 7, where the laurylamine exists as the ion, showed that 0.25 mg of laurylamine hydrochloride per liter (0.2 mg of free base per liter) showed "catching" of an air bubble. Hence the concentration of ion in a 25-mg-per-liter "solution" is probably sufficient to act as collector up to pH 10.6. The mechanism of collection is considered to be as follows:

TABLE 2—*Concentration of Laurylamine and Laurylammonium Ion in Solution at pH 8.5 to 13*

pH Value	$C_{12}H_{25}NH_2$, Mg per Liter	$C_{12}H_{25}NH_3^+$, Mg per Liter
8.5	0.2 (0.16)	20.8
10	0.2	0.8
10.6	0.2	0.2
11	0.2	8×10^{-2}
12	0.3	8×10^{-3}
13	0.2	8×10^{-4}

Up to pH 10 the laurylammonium ion is the collector; between pH 10 and 10.6 both ion and base act as collector and above pH 10.6 the free base is the effective agent. It is considered that above pH 10 any ion in solution is still preferentially adsorbed, but there is probably insufficient to form a sufficiently coherent film for an air bubble to adhere to. The free base is then adsorbed in the spaces on the mineral surface.

Taggart and Arbiter[18] state that if the collecting mechanism is of the type

$$R^+ + Y^- + MX \rightleftharpoons RX + M^+ + Y^-$$

where R^+ is the collector ammonium ion, Y^- the anion of the amine salt and M^+ and X^- are the cation and anion of the mineral, the reaction should obey qualitatively the mass action law; i.e. "coating and collection should be aided by the addition to the solution of either ion (R^+ or X^-) of the collector-coating compound, and hindered by the addition of the ions M^+ or Y^-." The statement that Y^- can affect the reaction is untenable, since it occurs on both sides of the equation and therefore cannot influence the equilibrium in any way.

This equation may be written for fluorite with laurylamine hydrochloride as collector as follows:

$$2LNH_3^+ + 2Cl^- + CaF_2 \rightleftharpoons 2LNH_3F + Ca^{++} + 2Cl^-$$

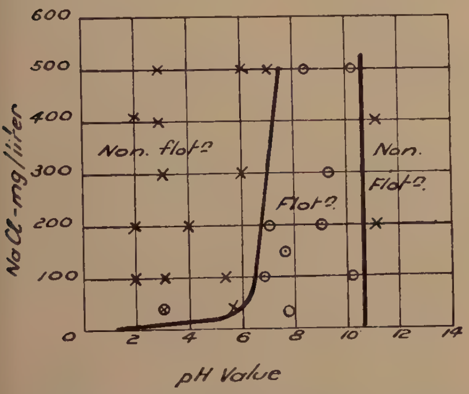

FIG 9—LAURYLAMINE HYDROCHLORIDE, 25 MG PER LITER, RELATIONSHIP BETWEEN SODIUM CHLORIDE CONCENTRATION AND pH VALUE—FLUORITE. TEMPERATURE, 35°C.

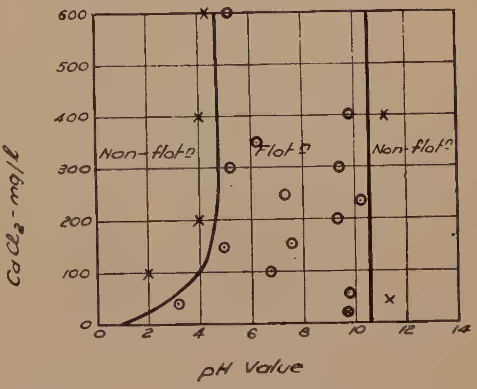

FIG 10—LAURYLAMINE HYDROCHLORIDE, 25 MG PER LITER. RELATIONSHIP BETWEEN CALCIUM CHLORIDE CONCENTRATION AND pH VALUE—FLUORITE. TEMPERATURE, 35°C.

Cylinder flotation tests with 25 mg laurylamine hydrochloride per liter as collector showed that calcium ion depresses fluorite in the acid region, there being no flotation below pH 5 (Fig 10). This depression is in accordance with Taggart's theory[18] but, according to this same theory, the flotation of calcite should also be depressed by the addition of calcium ion, whereas it is unaffected by concentrations of calcium ions up to 500 mg per liter. In general the presence in the system of an ion common to the mineral may affect the adsorption of collector in two ways: by competition with the collector ion for adsorption, or by precipitation of the collector ion. Calcium ion has no effect on the flotation of quartz. It could affect flotation only by competition with the laurylammonium ion at the mineral surface. Depressant effects by foreign cations are shown in Figs 9, 11 and 12. Depression by copper salts is due to soluble copper: Fig 11 shows that flotation becomes independent of the amount of copper chloride added at pH values above 6.6 because further additions merely result in precipitation of additional hydroxide or basic salts. (Although Britton[3] states that copper hydroxide precipitates at a pH value of 5.3, his solutions were very much more concentrated than those used in these tests, for which turbidity first appears at pH 6.6.)

Chromium acts as a depressant either in the form of the chromium ion or as "chromite ion" or peptized chromium hydroxide. At pH values up to 6 there is very little precipitate in solution although the solution is turbid. Between pH values of 6 and 11 there is a flocculent precipitate, which disappears between pH 11 and 12, either because of peptization of the chromium hydroxide or its solution as chromite. The conditions under which quartz does not float in the presence of chromium correspond with those under which chromium is soluble.

Taggart and Arbiter[18] say that depression of a mineral in the presence of an amine collector may be effected by addition of a small excess of a cation that forms with the mineral anion, a compound less soluble than the mineral itself. Calcium fluoride is more insoluble than any of the fluorides formed with the common cations, so this

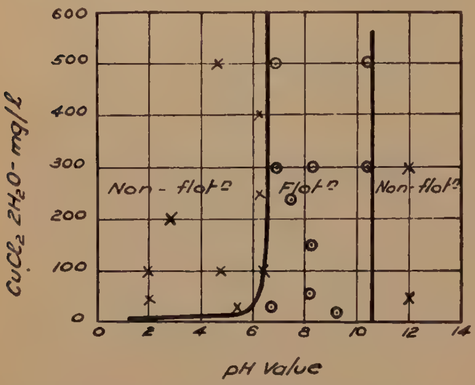

FIG 11—LAURYLAMINE HYDROCHLORIDE, 25 MG PER LITER. RELATIONSHIP BETWEEN CUPRIC CHLORIDE CONCENTRATION AND pH VALUE—FLUORITE. TEMPERATURE, 35°C.

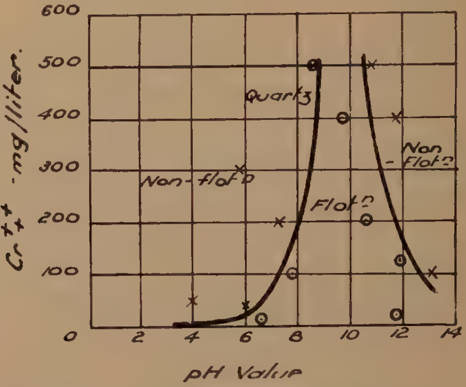

FIG 12—LAURYLAMINE HYDROCHLORIDE, 25 MG PER LITER. RELATIONSHIP BETWEEN CHROMIUM-ION CONCENTRATION AND pH VALUE—QUARTZ. TEMPERATURE, 35°C.

theory of depression cannot be tested for fluorite. However, results show that to be a depressant a cation need not necessarily form a fluoride more insoluble than fluorite, for sodium chloride depresses fluorite with laurylamine as collector and sodium fluoride is much more soluble than calcium fluoride. Also copper fluoride is more soluble than calcium fluoride and yet copper depresses fluorite (Table 3). Other data, summarized in Table 3, are in agreement with Taggart's theory, except possibly for the lack of depression of quartz by calcium ion.

Taggart and Arbiter state[18] that the amount of a given cationic collector necessary for flotation may be decreased sharply by adding to the pulp the anion of the mineral. Our results show that sodium fluoride does not activate fluorite, and may even depress it. In cylinder flotation tests the addition of 6 mg of laurylamine hydrochloride per liter results in good flotation at 35°C. According to

TABLE 3—*Effect of* Na^+, Cu^{++}, Cr^{+++} *on Flotation of Fluorite, Calcite, Quartz, with Laurylamine Hydrochloride as Collector*

Mineral	Laurylamine HCl = 25 Mg per Liter		
	Added Ion	Effect on Flotation	Solubility of Metal Salt, Grams per 100 Ml Water[9]
Fluorite	Na^+	Depression below pH 7	NaF = 4.22 at 18°C
Fluorite	Cu^{++}	Depression below pH 6.6	$CuF_2 \cdot 2H_2O$, slightly soluble CuF_2, insoluble
Fluorite	Ca^{++}	Depression below pH 5	CaF_2, insoluble
Fluorite	Cr^{+++}	No depression	Soluble
Calcite	Na^+	No depression	Na_2CO_3, 7.1 at 0°C
Calcite	Cu^{++}	No depression	Copper carbonate, insoluble (cold)
Calcite	Ca^{++}	No depression	
Calcite	Cr^{+++}	No depression	Chromium carbonate, veryslightly soluble
Quartz	Na^+	No depression	Sodium silicates, soluble
Quartz	Cu^{++}	Depression from pH > 13 to pH 11	Copper silicates, insoluble
Quartz	Ca^{++}	No depression	Calcium silicates, insoluble
Quartz	Cr^{+++}	Depression in both alkaline and acid regions	Chromium silicates, insoluble

the hypothesis, sodium fluoride in solution should give flotation with concentrations of laurylamine hydrochloride less than 6 mg per liter; e.g., 4 mg per liter. However, even with 6 mg per liter there is no flotation in the presence of 50 mg NaF per liter. These tests were made at pH 7.5 to 8, where sodium ion causes no depression (Table 3 and Fig 9) and the collector exists as the ion.

Although the laurylammonium ion is cationic and so affected by cations in solution, the free base may be affected by anions or cations in solution, and so anions may affect flotation with laurylamine above pH 10.6.

Flotation-machine Tests with Igepon T and Aerosols OT and MA

For all flotation tests, ore from Pine Mountain, Victoria, assaying 83 to 89 pct fluorite, 8 to 12 pct quartz, about 1 pct calcite and small amounts (less than 1 pct) of galena and an iron sulphide was used. For acid-grade fluorite, the quartz content must be reduced below 1 pct. The ore was dry-ground to minus 10-mesh and test samples stage-ground in water to minus 52-mesh (B.S.S.) in a porcelain pebble mill or in a steel ball mill. The material was floated in 500-gram samples in a 500-gram (3-liter) Denver subaeration machine. The mineral was conditioned with the collector for one minute before floating and no other reagent was added, Igepon and Aerosol providing good froths. The pulp was kept at 35 ± 1°C, and the pH of the pulp was checked by the glass electrode, the floats being continued until the froth was free of mineral.

The low percentage of calcite in the ore (1 pct) enabled the pH of the pulp to be lowered to 2 by the addition of the theoretical amount of acid but the value slowly rose during the float, through the solution of this calcite, and the value after the float was 2.5 to 3. These are the pH values meant when it is stated in the text that the mineral was floated at pH 2. The results of these tests are represented graphically in Figs 14 to 16, and from them the following conclusions can be drawn:

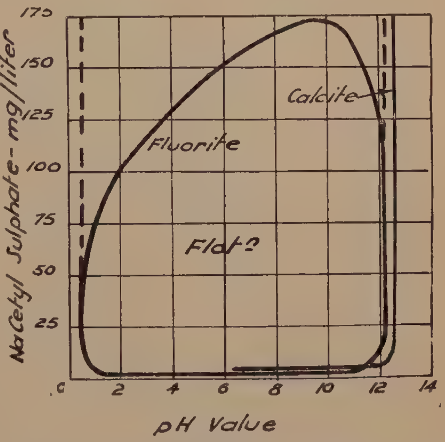

FIG 13—RELATIONSHIP BETWEEN SODIUM CETYL SULPHATE CONCENTRATION AND pH VALUE—FLUORITE, CALCITE. TEMPERATURE, 35°C.

1. The percentage of quartz floating in the concentrate increases markedly as the pH value increases (Fig 14). The quartz in the concentrate is due partly to mechanical carry-over in the froth and this will increase with pH value as the froth becomes more voluminous. It is considered that flotation of quartz is due to activation by lead (from galena in the ore) or iron (from the ball mill) during grinding. During grinding the calcite in the ore keeps the pH of the pulp above 5, but it has been found that iron activates quartz during grinding in a pulp with a pH as high as 10.[8] Lead and iron activation will be partly removed at low pH values by competition with hydrogen ion, the deactivation being greatest at the lowest pH value, and so the amount of activated quartz floating is a minimum at low pH values.

2. The recovery of fluorite also increases with pH value (Fig 14). The results

given are for rougher floats only. The recovery of fluorite with Aerosol OT is above 85 pct at pH 2 and so cannot increase appreciably. The recovery with Igepon T is much poorer throughout, but an increase in the amount of Igepon used helps the recovery (Fig 15). The results again are for rougher floats only.

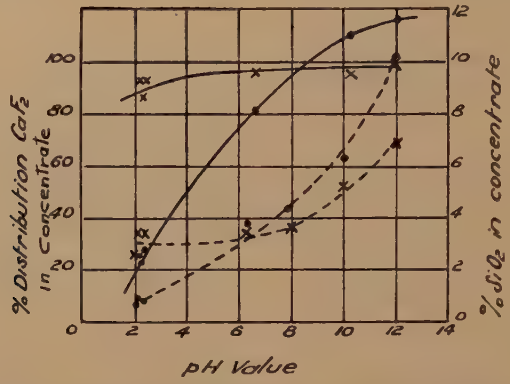

Fig 14—Aerosol OT, 25 mg per liter; Igepon T, 25 mg per liter; relationship between pH value and percentage distribution of fluorite and percentage of silica in concentrate.

One minute conditioning, 35°C, pH 2.
Aerosol OT, full line: Igepon T, dashed line.
Dot indicates percentage SiO_2; cross, percentage distribution CaF_2.

3. The amount of quartz floating also increases with increase in Igepon concentration (Fig 15).

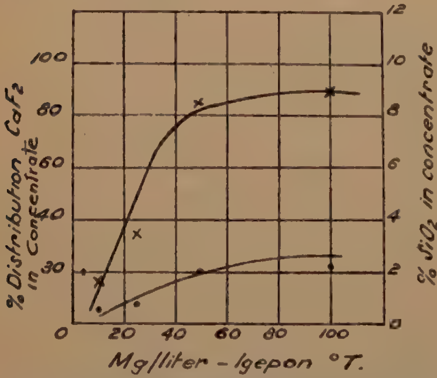

Fig 15—Relationship between Igepon T concentration and percentage distribution of fluorite and percentage of silica in concentrate.

One minute conditioning, 35°C, pH 2. Dot indicates percentage SiO_2; cross, percentage distribution CaF_2.

4. Both the percentage of quartz in the concentrate and the percentage of fluorite floating in the rougher float increase with temperature increase from 15° to 35°C (Fig 16). Higher temperatures have not been investigated and the optimum temperature seems to be 30° to 35°C.

5. Aerosol OT is a much better collector for fluorite than is Igepon T. Although

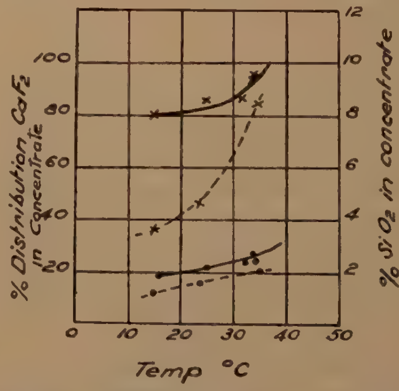

Fig 16—Igepon T, 50 mg per liter; Aerosol OT, 25 mg per liter. Relationship between temperature and percentage distribution of fluorite and percentage of silica in concentrate.

One minute conditioning, pH 2.
Aerosol OT, full line; Igepon T, dashed line.
Dot indicates percentage SiO_2; cross, percentage distribution CaF_2.

the amount of quartz floating is slightly higher with Aerosol, this can be satisfactorily reduced in subsequent cleaner floats. Tests with Aerosol MA gave results very similar to those with Aerosol OT.

6. Variations in conditioning time were also investigated. An increase in conditioning time resulted in an increase in fluorite recovery and in the percentage of quartz floating.

By subjecting the rougher concentrates to cleaning, the quartz content was reduced to less than 1 pct, but the recovery was poor. The results in Table 4 were obtained by refloating rougher concentrates without return of cleaner tailings. Under operating conditions over-all recoveries would be much higher.

TABLE 4—*Flotation of Fluorite with Igepon T and Aerosol OT as Collectors at* 35°C

Reagent	pH of Pulp	Product, Concentrates	SiO_2 in Conc., Pct	CaF_2 in Conc., Pct	Distn. CaF_2, Pct
Igepon T, 0.6 lb per ton	2.0–2.5	Rougher	2.00	97.7	86.5
	2.0–2.2	Cleaner	0.87	98.9	58.2
Igepon T, 1.2 lb per ton	2.0–2.4	Rougher	2.16	95.7	88.1
	2.0–2.2	Cleaner	0.76	97.5	51.4
Aerosol OT, 0.3 lb per ton	2.0–2.6	Rougher	2.48	94.7	92.4
	2.0–2.3	Cleaner	0.55	97.4	47.4

Typical results with Aerosol OT and Igepon T are given in Table 4 for floats at 35°C.

Sodium Cetyl Sulphate

Sodium cetyl sulphate in solution yields the anion $C_{16}H_{33}SO_4$, which is the collector. It is adsorbed on fluorite and competes with the fluoride ions for a place in the surface of the lattice. Hence fluoride ions would be expected to have a depressant effect and calcium ions an activating effect, which is the reverse of what happens when the collector is cationic. Fig 13 shows the behavior of fluorite and calcite with sodium cetyl sulphate as collector. The reagent is not a collector for quartz. Flotation of fluorite ceases over the entire pH range when the sodium cetyl sulphate concentration reaches 175 mg per liter and at lower collector concentrations it is affected as the ionic concentrations are increased by addition of acid or alkali. That there is an adsorbed film on the mineral when flotation ceases in solution is shown by removing the mineral into water of the same pH value as the solution when excellent flotation or air-bubble contact is obtained from pH values of 1 to 12. Calcite shows no cessation of flotation at 175 mg per liter (Fig 13).

If an anionic collector reacts with a mineral according to Taggart's equation

$$MX + Y^+R^- \rightarrow MR^- + Y^+ + X^-$$

the compound formed at the mineral surface with sodium cetyl sulphate should be the same for calcite and fluorite; viz., calcium cetyl sulphate; and one must conclude that the flotation behavior of both minerals and, indeed, all calcium minerals with anionic collectors, should be closely related. However, the flotation behaviors of fluorite and calcite often differ appreciably, as shown by Fig 3 and by the fact that the Aerosols do not float calcite but float fluorite well (Figs 1 and 2).

TESTS IN FLOTATION MACHINE

Laboratory tests having given promising results, the work with Pine Mountain fluorite was extended to flotation tests in a 500-gram Denver subaeration machine and then in a similar 2000-gram machine. The ore was stage-ground in water to minus 52-mesh (B.S.S.) in a porcelain pebble mill or a steel mill containing flints.

The floats were done under the following conditions:

1. Pure sodium cetyl sulphate as collector:

a. Ore ground in porcelain ball mill, floated in 500-gram cell.

b. Ore ground in porcelain mill, floated in 2000-gram cell.

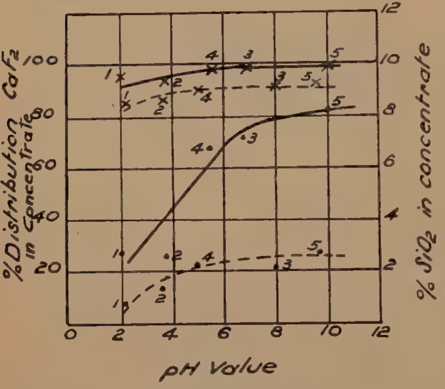

FIG 17—SODIUM CETYL SULPHATE, 25 MG PER LITER (0.25 LB PER TON). RELATIONSHIP BETWEEN pH VALUE AND PERCENTAGE DISTRIBUTION OF FLUORITE AND PERCENTAGE OF SILICA IN CONCENTRATE.

One minute conditioning, 35°C.

Rougher float, full line; cleaner, dashed line.

Dot indicates percentage SiO_2; cross, percentage distribution CaF_2.

Results from corresponding rougher and cleaner floats given same number.

c. Ore ground in steel mill, floated in 2000-gram cell.

2. Commercial sodium cetyl sulphate as collector:

a. Ore ground in steel or porcelain mill, floated in 500-gram cell.

b. Ore ground in steel mill, floated in 2000-gram cell.

The ore was conditioned with the collector in the flotation machine prior to the rougher float, and the rougher concentrate was repulped with water for the cleaner float, with the addition of acid (H_2SO_4) or alkali (NaOH) to adjust the pH value. The floats were all batch, with no return of flotation liquor or cleaner tailings. From the floats with pure reagent can be seen the effect of pH value (Fig 17) and temperature (Fig 18) of the pulp on the percentage quartz and the percentage distribution of fluorite in the concentrate.

Corresponding rougher and cleaner concentrates are denoted by the same number. The recovery of fluorite increases with pH value of the pulp (Fig 17), although the recoveries are good with 25 mg of sodium cetyl sulphate per liter (0.25 lb per ton) as collector, even at pH 2. In the cleaner float the recovery is also

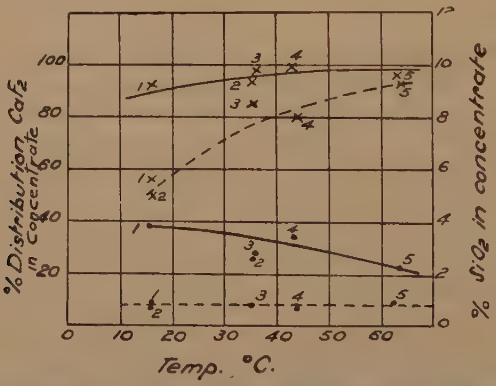

FIG 18—SODIUM CETYL SULPHATE, 25 MG PER LITER. RELATIONSHIP BETWEEN TEMPERATURE AND PERCENTAGE DISTRIBUTION OF FLUORITE AND PERCENTAGE OF SILICA IN CONCENTRATE.

One minute conditioning pH 2.

Rougher float, full line: cleaner, dashed line.

Dot indicates percentage SiO_2; cross, percentage distribution CaF_2.

Results from corresponding rougher and cleaner floats given same number.

slightly less at lower pH values, although the collector film is already on the mineral.

The percentage of quartz in the concentrate also increases markedly with pH value, particularly in the rougher floats, probably because the activation of quartz is less at low pH values.

Increase in temperature from room (16°C) to 65°C (Fig 18) increases the recovery of fluorite: this effect is more marked in the cleaner float. The percentage of quartz in the concentrate is not greatly affected by temperature. (An apparent exception—No. 1, Table 8—probably is due to the unusually high percentage of quartz in the head sample.)

Sodium cetyl sulphate 50 mg per liter (0.5 lb per ton) gives excellent recoveries of fluorite in the cleaner concentrate, but also floats more quartz. Table 5 compares the results with 10, 25 and 50 mg per liter at 35°C.

Two commercial products containing approximately 16 pct sodium cetyl sulphate were used as collectors. They were poor substitutes for the pure material, mainly because recoveries were poorer, even when amounts were used to give true concentrations of 25 mg of sodium cetyl sulphate per liter or more. One major impurity, sodium sulphate, had no detrimental effect on flotation of fluorite by pure sodium cetyl sulphate. Fig 19 shows typical results obtained with one of these reagents.

Conclusions

From the point of view of obtaining from Pine Mountain fluorite ore a fluorite concentrate containing less than 1 pct of quartz and with reasonable recoveries of fluorite, the following facts emerge from the data so far presented:

1. Of the anionic reagents investigated sodium cetyl sulphate is the best collector, with Aerosols OT or MA better than Igepon T.

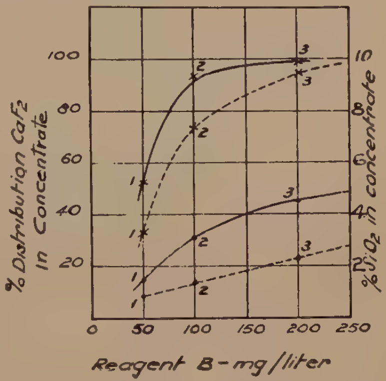

FIG 19—RELATIONSHIP BETWEEN REAGENT B CONCENTRATION AND PERCENTAGE DISTRIBUTION OF FLUORITE AND PERCENTAGE OF SILICA IN CONCENTRATE.

Natural pH (6–7), 35°C. One minute conditioning.

Rougher float, full line; cleaner, dashed line.

Dot indicates percentage SiO_2; cross, percentage distribution CaF_2.

Results from corresponding rougher and cleaner floats given same number.

TABLE 5—*Flotation of Fluorite by Sodium Cetyl Sulphate*

pH of Float	Concentrated Sodium Cetyl Sulphate, Mg. per Liter	Product, Concentrate	CaF_2 (Analysis), Pct	CaF_2, Distn., Pct	SiO_2, (Analysis), Pct	SiO_2, Distn., Pct
2–3	10	Rougher	95.8	14.8	2.33	2.4
2–3	25	Rougher	94.6	96.8	2.70	29.3
		Cleaner	96.4	84.0	0.85	7.8
2–3	25	Rougher	93.7	98.5	3.35	33.0
		Cleaner	96.1	79.8	0.73	5.6
2–3	50	Rougher	95.5	97.7	2.17	20.7
		Cleaner	96.3	96.8	1.35	12.7
7–8 (natural pH of pulp)	10	Rougher	91.6	58.1	7.53	32.7
7–8 (nat.)	25	Rougher	90.4	99.0	7.09	68.1
		Cleaner	95.9	89.6	2.17	17.9
7–8 (nat.)	50	Rougher	88.3	99.7	8.63	84.9
		Cleaner	93.3	98.5	4.04	37.1

2. The two commercial sodium cetyl sulphate samples used are poor substitutes for the pure reagent.

3. With sodium cetyl sulphate the optimum pH value for flotation is 2 to 5, preferably 2 to 3 to keep quartz down.

4. With sodium cetyl sulphate the optimum temperature for flotation is above 30°C, 30° to 40° being satisfactory.

Table 6—*Flotation of Fluorite by Sodium Cetyl Sulphate*

Collector	pH Value	Temperature, Deg C	Product[a]	Wt. Pct	CaF₂, Pct	CaF₂ Distn., Pct	SiO₂, Pct	SiO₂ Distn., Pct
Pure sodium cetyl sulphate, 25 mg per liter	2–2.5	16	Cleaner c.	49.6	96.6	57.2	0.86	3.3
			Cleaner t.	33.4	89.8	35.9	8.09	20.4
			Rougher c.	83.0	93.9	93.1	3.77	23.7
			Rougher t.	17.0	33.8	6.9	59.2	76.3
			Head	100.0	83.6	100.0	13.2	100.0
Pure sodium cetyl sulphate, 25 mg per liter	2–3	35	Cleaner c.	76.8	96.4	84.0	0.85	7.8
			Cleaner t.	13.4	84.2	12.8	13.4	21.5
			Rougher c.	90.2	94.6	96.8	2.70	29.3
			Rougher t.	9.8	29.0	3.2	59.9	70.7
			Head	100.0	88.1	100.0	8.31	100.0
Pure sodium cetyl sulphate, 25 mg per liter	2–3	44	Cleaner c.	71.7	96.1	79.9	0.73	6.47
			Cleaner t.	19.2	84.6	18.7	13.2	31.0
			Rougher c.	90.9	93.6	98.6	3.36	37.4
			Rougher t.	9.1	13.8	1.4	56.2	62.6
			Head	100.0	86.4	100.0	8.17	100.0
Pure sodium cetyl sulphate, 25 mg per liter	2–2.5	64	Cleaner c.	82.8	96.5	93.0	0.90	6.9
			Cleaner t.	4.6	69.5	3.7	26.3	11.1
			Rougher c.	87.4	95.1	96.7	2.24	18.0
			Rougher t.	12.6	22.8	3.3	70.9	82.0
			Head	100.0	86.0	100.0	10.9	100.0
Sodium cetyl sulphate, commercial A = 1000 mg per liter	2–2.5	36	Cleaner c.	76.3	96.5	83.2	0.91	7.8
			Cleaner t.	3.1	81.3	2.8	13.9	4.9
			Rougher c.	79.4	95.9	86.0	1.42	12.7
			Rougher t.	20.6	60.0	14.0	37.3	87.3
			Head	100.0	88.5	100.0	8.80	100.0
Sodium cetyl sulphate, commercial B = 100 mg per liter	5–6	35	Cleaner c.	55.1	97.3	60.3	0.80	5.0
			Cleaner t.	17.7	93.1	18.5	4.85	9.8
			Rougher c.	72.8	96.3	78.8	1.79	14.8
			Rougher t.	27.2	69.4	21.2	27.4	85.2
			Head	100.0	89.0	100.0	8.75	100.0
Commercial B (purified) = 100 mg per liter	7.5–8	35	Cleaner c.	74.0	97.0	81.8	1.45	11.1
			Cleaner t.	12.0	89.5	12.2	8.00	9.9
			Rougher c.	86.0	96.0	94.0	2.36	21.0
			Rougher t.	14.0	38.0	6.0	54.5	79.0
			Head	100.0	87.8	100.0	9.66	100.0

[a] c indicates concentrates; t, tailings.

Table 6 shows typical results illustrating these conclusions.

Adsorption of Sodium Cetyl Sulphate

The adsorption of sodium cetyl sulphate on fluorite was determined by titration of the residual cetyl sulphate in solution, by the method of Auerbach[1] as modified by Rogers et al.[15]

The fluorite was hand-picked material from Chillagoe, Queensland, containing 96 pct fluorite. It was wet-ground in a porcelain mill to obtain a minus 325-mesh (Tyler) fraction. The method was as follows:

A measured volume of the fluorite suspension (5 ml) was shaken in a standard flask with a known concentration of cetyl sulphate and the flask was kept at 35°C for a definite time. The liquid was then separated from the fluorite in a centrifuge, its pH value was measured and the residual cetyl sulphate estimated by titration against cetyl trimethyl ammonium bromide.[15] The excess cetyl trimethyl ammonium bromide reacts with bromphenol blue to form a blue compound soluble in ethylene dichloride and the color may be matched visually against a known blank. Filtration of the pulp through filter paper or sintered glass is not permissible, since both adsorb cetyl sulphate. The fluorite after centrifuging was filtered, dried and weighed. The concentration of the stock solution of sodium cetyl sulphate was 0.250 mg per liter, which is slightly less than the solubility of the compound at 35°C.

Table 7 shows that adsorption increases with time of contact between the mineral and cetyl sulphate.

TABLE 7—*Effect on Adsorption and Contact Angle of Time of Contact between Fluorite and Sodium Cetyl Sulphate*

35°C at natural pH value 6.4 to 6.7. 10 mg sodium cetyl sulphate added for adsorption tests, 25 mg per liter sodium cetyl sulphate for contact angle

Time in Thermostat	Contact Angle, Deg	Sodium Cetyl Sulphate Adsorbed, Mg per Gram Fluorite	Residual Sodium Cetyl Sulphate Conc., Mg per Liter in Adsorption Tests
5 min.	76	2.24	34.0
10 min.	78		
15 min.		2.57	38.0
20 min.	83		
30 min.	85	2.90	35.0
1 hr	86	4.18	36.4
2 hr	86	4.52	35.9
hr		4.38	35.8
5 hr		6.25	34.5

The contact angles that develop under similar conditions are also shown. Constant contact angles are reached in about 30 min. but adsorption continues for several hours. At pH values of 1 and 9 to 11 (Table 8) adsorption equilibrium is reached more quickly, for the amount adsorbed in 30 min. at these values is close to that adsorbed in 5 hr at a pH value of 6.5 (Table 7).

To obtain correlation with contact and flotation tests, adsorption usually was determined after 30 min., although equilibrium may not have been reached.

Table 8 shows the effect on adsorption of the pH value of the pulp.

TABLE 8—*Effect of* pH *Value of Pulp on Adsorption of Sodium Cetyl Sulphate on Fluorite*

30 minutes at 35°C

pH Value of Supernatant Liquid	Sodium Cetyl Sulphate Added, Mg	Sodium Cetyl Sulphate Adsorbed per Gram Fluorite, Mg	Sodium Cetyl Sulphate Residual Conc., Mg per Liter
1	10	5.71	18.8
1	10	5.74	16.0
2	10	2.87	16.1
2	10	2.74	15.0
6.7	10	2.66	20.1
6.7	10	2.88	13.2
6.7	10	2.86	11.7
6.9	10	2.79	14.1
9.6	10	5.63	19.8
9.4	10	5.91	14.0
9.1	10	5.01	10.3
10.4	10	5.90	13.7
10.1	10	5.61	15.6
11	10	5.59	12.4
11	10	5.90	13.7
11	10	5.61	15.6
12	2.5	0.32	15.4
12	2.5	0.34	15.3

From Fig 13 one would expect adsorption to be greatest at pH values of 6 to 11 and to fall rapidly above pH 12 and below pH 1. It is therefore surprising that the amount adsorbed is so high at pH 1. Possibly the added acid hastens adsorption and the added alkali has a similar effect at high pH values. In neutral solutions this greater amount of cetyl sulphate is adsorbed only after several hours.

Fig 13 shows that 175 mg of the collector per liter is sufficient to prevent contact of an air bubble in the collector solution. To determine whether this cessation of contact corresponds to increased adsorption, and perhaps to the formation of a partial or complete double layer, adsorption tests were made in solutions of higher concentrations of cetyl sulphate. Table 9 shows that there is increased adsorption, although in no case was a residual cetyl sulphate concentration as high as 175 mg per liter reached.

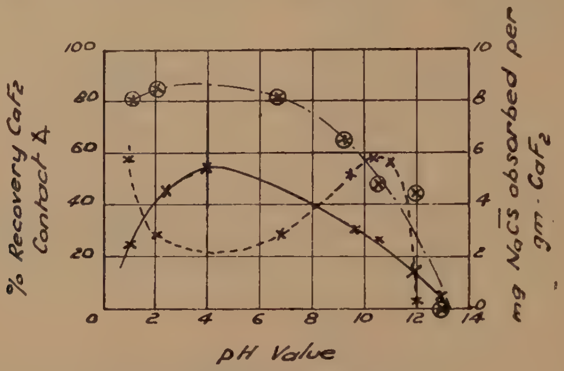

FIG 20—SODIUM CETYL SULPHATE. 35°C.
Relationship between pH value and:
1. Percentage distribution of fluorite.
2. Contact angle.
3. Milligrams sodium cetyl sulphate adsorbed per gram of fluorite.
Full line indicates percentage recovery of fluorite; dot and dash line, contact angle; dashed line, milligrams sodium cetyl sulphate adsorbed per gram fluorite.

TABLE 9—*Effect of Concentration of Sodium Cetyl Sulphate Added on Adsorption*
30 minutes at 35°C

Sodium Cetyl Sulphate Added, Mg	pH Value	Sodium Cetyl Sulphate Adsorbed, Mg per Gram CaF_2	Residual Sodium Cetyl Sulphate Conc., Mg per Liter
10	2	2.80	15.5
20	2	9.39	58.0
10	6.7	2.88	13.2
20	6.7	3.52	134
10	9.4	5.91	140
20	9.4	10.4	50.0

The adsorption at pH 2 is increased more than threefold and at pH 9 nearly twice, but at the natural pH there is very little increase in adsorption when the amount of sodium cetyl sulphate added is doubled. This curious result may be connected with the effect of inorganic ions on adsorption mentioned previously.

Fig 20 shows the effect of pH value on adsorption, contact angle and recovery of fluorite in a flotation machine. The floats were down with −52+100-mesh (B.S.S.) fractions of Pine Mountain fluorite, obtained from ore that was ground to minus 52-mesh, sized and sampled. Each sample of 460 grams was floated in a 500-gram (3-liter) Denver machine at 35°C. Pine oil (20 mg per liter) was first added to the pulp and any floating material was removed, then 25 mg of pure sodium cetyl sulphate per liter was added and the pulp conditioned for 2 min. before flotation. Each float was continued for 3 min., although some were over in a shorter time. Samples of the pulp were taken at the end of the conditioning period and at the end of the float and were centrifuged as soon as they were taken, and the sodium cetyl sulphate content was determined.

Table 10 shows the amount of sodium cetyl sulphate adsorbed in each test. Of the 460-gram sample, approximately 400 grams was fluorite of particle size −52+100-mesh (B.S.S.). It thus has an area perhaps 28 times smaller* than in the adsorption tests. Allowing for the difference in particle size, the amount of adsorption that corresponds to the amount determined in the adsorption tests has been calculated and is given in column 6 of Table 10.

TABLE 10—*Amount of Sodium Sulphate Adsorbed in Flotation Tests on Sized Mineral*
25 Mg sodium cetyl sulphate per liter added (75 mg)

pH of Pulp	Sodium Cetyl Sulphate in Pulp Mg per Liter		Sodium Cetyl Sulphate Adsorbed, Mg		Sodium Cetyl Sulphate Adsorbed Under Similar Conditions Calculated from Adsorption Tests, Mg	Recovery of Fluorite in Float, Pct
	At End of Conditioning	At End of Float	At End of Conditioning	At End of Float		
1	0.2		74	75	82	23.4
2–2.5	0.9	0.2	72	74	39	45.5
6.1–6.4	1.9	0.2	69	74	39	54.5
10.0–9.0	2.8	1.9	67	69	82	30.0
11–10.3	3.5	1.9	64	69	82	26.6
12–11.1	3.3	1.4	65	71	5	16.3
13	6.9	6.6	54	55		5.4

At pH values of 2 to 6, adsorption has proceeded further in 2 min. in the flotation tests than in 30 min. in the adsorption tests, presumably because of better agitation.

Even at the most favorable pH values, the recovery of fluorite in Fig 20 does not reach 60 pct. An increase in conditioning period from 2 to 30 min. produced no increase in recovery. The significant factor is probably particle size, for a sizing analysis of the concentrate and tailing of the test with 10 mg of sodium cetyl sulphate per liter at a pH value of 7 to 8 (Table 5) gave the distribution:

B.S.S.	Concentrate, Pct	Tailing, Pct
−52 + 100-mesh	19	81
−100 + 200-mesh	31	69
−200-mesh	61	39

From Fig 20 it seems that at pH values above 7 and below 2 more sodium cetyl sulphate is adsorbed per gram of fluorite, and that this larger amount is necessary for flotation to be as good as with smaller amounts within the range 2 to 7. That the amount of sodium cetyl sulphate adsorbed at pH values of 2 to 7 is sufficient for flotation is also borne out by the contact-angle curve, in which the contact angle is a maximum over this pH range.

Summary

The flotation of fluorite and the gangue minerals calcite and quartz by various types of collectors has been investigated. Some of these reagents, in particular sodium cetyl sulphate, may be worthy of attention as alternatives to oleic acid for floating fluorite; for this reason tests with commercial samples of sodium cetyl sulphate were made. The major advantage in the use of sodium cetyl sulphate instead of oleic acid would be that considerably lower temperatures could be used.

Table 13 summarizes the flotation behavior of fluorite, calcite and quartz with the collectors investigated.

Fluorite and calcite are floated by both anionic and cationic collectors and quartz only by cationic ones.

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* The particle size of the fluorite in the flotation tests is between 0.295 mm (52-mesh B.S.S.) and 0.152 mm (100-mesh B.S.S.) and in the adsorption tests is less than 0.044 mm (325-mesh Tyler). Hence the surface area in the flotation tests is 3.4 to 6.6 (average 5) times *smaller* than in the adsorption tests. Actually all the material in the adsorption tests passes 325-mesh. Thus the true conversion factor must be considerably greater than 5. Assuming it to be 28, there is reasonable agreement between the adsorption and flotation tests (Table 10).

Successful separations of fluorite from quartz have been obtained with sodium cetyl sulphate as collector at low pH values.

TABLE 13.—*Flotation of Fluorite, Calcite and Quartz*

Reagent	Fluorite	Calcite	Quartz
Aerosol OT....	Flotation	No flotation	No flotation
Aerosol MA...	Flotation	No flotation	No flotation
Igepon T......	Flotation	Flotation	No flotation
Ninol 737.....	Flotation	Flotation	Flotation
Sapamine M. S.	Flotation	No flotation	Flotation
Cetyl trimethyl ammonium bromide.....	Flotation	No flotation	Flotation
Cetylamine hydrochloride..	Flotation	Flotation	Flotation
Laurylamine hydrochloride	Flotation	Flotation	Flotation
Sodium cetyl sulphate.....	Flotation	Flotation	No flotation

The amount of sodium cetyl sulphate adsorbed on fluorite has been measured under different conditions of temperature, pH value and time.

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The Application of Xanthates to Flotation

By Norman Weiss,* Member AIME

(New York Meeting, March 1947)

Most papers on xanthate have dealt with principles rather than practice. On the assumption that many millmen are interested in knowing where and in what manner the xanthates are being used in mills other than their own, this paper will summarize the results of a recent survey of North American plants. A brief background story and a description of the xanthates will be followed with the results of that survey; then a few observations will be made on storage, feeding, and ore testing.

The first xanthate was made in 1822 by Zeise of the University of Copenhagen. With commendable scientific detachment (or perhaps a bad cold) Zeise ignored a more pervading quality of the substance and named it for its color, after the Greek word "xanthos," meaning yellow. During the first 80 years of its recorded existence xanthate found use only in small quantity as an insecticide; after the turn of the century there was some further demand for it in the vulcanization of rubber. It was not until 1922, by coincidence the centennial of Zeise's discovery, that C. H. Keller of San Francisco, seeking sulphidizing agents, recalled xanthate in connection with his experience years earlier when he had seen this material used in Europe to combat grape phylloxera. His successful tests with the new flotation reagent culminated in the basic xanthate patent (U. S. 1,554,216) in 1925.

In the minds of many metallurgists the introduction of xanthate as a flotation reagent is identified with the beginnings of alkaline circuits, the first use of chemical collectors, and the beginning of selective flotation. Actually, these preceded xanthate by a number of years. The alkaline circuit in lead-zinc flotation was introduced by Lyster as early as 1912. In 1922 Sheridan and Griswold made their momentous discovery (Timber Butte) that alkali cyanides and zinc sulphate effectively inhibited the flotation of pyrite and sphalerite in pulps made alkaline with sodium carbonate or bicarbonate. The use of lime to retard pyrite in copper flotation began within two years after Ralston directed attention to its possibilities in 1917.

These were the beginnings of modern selective flotation, which was practiced successfully on copper-iron ores at Nacozari in 1922, and at Inspiration, Cananea, and others only shortly afterward. In the lead-zinc mills, selective flotation was of even greater economic importance at that time. The Sunnyside plant of U. S. Smelting, Refining, and Mining Co. used the process earlier than 1920, and Consolidated, Timber Butte, and others soon followed.

As a chemical collector xanthate was relatively a latecomer, for the Corliss patent of 1917 covered the first of these, alpha-naphthylamine, and the Perkins patent of 1919 the second, thiocarbanilid. Both of these were in extensive use before Keller's xanthate made its commercial debut in 1923.

There is general agreement, nonetheless, that the introduction of xanthate was

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momentous and opportune, for as an adjunct to the other important developments of that "golden age of flotation" it frequently supplanted frustration with success. Those who struggled with selective flotation during its infancy will recall the skepticism and then the profound relief with which the first of the xanthates was welcomed. Welcomed they were indeed, for in a very short time the xanthates attained prominence as collectors for the sulphide minerals, for some of the oxidized minerals of copper and lead, for such elemental metals as gold, silver, copper, and such nonmetals as graphite and sulphur. While this accomplishment was of prime importance there was also a secondary role for the new reagent; to the research scientists who sought a complete theory of flotation it afforded a precise and convenient tool. Consequently, in much of the valuable work on the theory of flotation that has been published during the past 20 years xanthate collectors have been employed, principally because they are specifically suited to the control and measurement required in such investigations, but also because their wide use has made them a logical standard of comparison.

TABLE 1—*The Commercial Xanthates*

Xanthate	Common Name	Formula	Molecular Weight	Number of Carbons[a]
Potassium ethyl	Potassium xanthate	$C_2H_5OCS_2K$	160.3	2
Sodium ethyl	Sodium xanthate	$C_2H_5OCS_2Na$	144.2	2
Potassium isopropyl	Isopropyl xanthate	$C_3H_7OCS_2K$	174.3	3
Sodium isopropyl	Isopropyl xanthate	$C_3H_7OCS_2Na$	158.2	3
Potassium sec-butyl	Butyl xanthate	$C_4H_9OCS_2K$	188.3	4
Sodium sec-butyl	Butyl xanthate	$C_4H_9OCS_2Na$	172.2	4
Potassium amyl	Amyl xanthate	$(C_2H_5)_2CHOCS_2K$	202.3	5
Potassium pentasol	Pentasol xanthate	$C_5H_{11}OCS_2K$	202.3	5

[a] Number of carbon atoms in the alkyl group.

DESCRIPTION OF THE XANTHATES

Xanthates are alkyl sulphur derivatives of carbonic acid, known as dithiocarbonates. Strictly, a xanthate is any salt of xanthic acid, but the xanthate flotation reagents are specifically the reaction products of carbon disulphide, an aliphatic alcohol, and either caustic soda or potash. The general formula is *ROCSSM*, in which *R* stands for an alkyl hydrocarbon radical and *M* for an alkali metal. Many xanthates have been compounded on the basis of this reaction, some of them only laboratory curiosities, others possessing desirable qualities in varying degree, but eventually the number of alcohols (including their homologues) was narrowed to four: the ethyl, isopropyl, butyl, and amyl.

The first xanthate on the market, the potassium ethyl xanthate initially used at Anaconda, was a relatively pure material but its manufacturing process resulted in a mother-liquor by-product. Improvements in the methods of manufacture soon resulted in the elimination of this by-product and a dry material of high purity and stability was produced. Then followed the sodium ethyl xanthate, a number of potassium amyl xanthates, several butyl and isopropyl xanthates, and finally a hexyl xanthate. The last was discontinued after only a short period of manufacture. At present only the eight xanthates listed in Table 1 are available in commercial quantities. The number of carbon atoms in the alcohol radical, shown in the last column of the tabulation, is sometimes considered a yardstick of collector power; for this reason the xanthates (and other collectors too) are often thus designated, the ethyl xanthates being known as C_2, the butyl as C_4, etc. In later discussion these symbols will be used. Attention is directed to the fact that Table 1 does not include all xanthate reagents.

The commercial xanthates are yellow to yellowish gray powders or pellets, completely and readily soluble in water, and possessing a characteristic odor owing to the presence of very minute amounts of mercaptans. They are noncorrosive, stable, uniform, and of relatively high purity. As regards their qualities as flotation reagents they are substantially nonfrothing and quick-acting, the former being highly desirable because frothing and collecting are important functions that require separate control in accordance with individual ore requirements and periodical variations in the circuit, the latter because xanthate quickly diffuses through the liquid phase and thus becomes immediately effective without recourse to pumps, conditioners, or emulsifiers. Further, since froth "flocculation" is often considered a special attribute of the xanthates, a word on that subject is indicated.

FLOCCULATION

When a heavily-laden froth is broken down by a spray of water the mineral particles appear to cling together and many observers have assumed this "flocculation" to be an indication and even a requisite of good flotation. This belief is only partly substantiated by the facts. There is but one essential condition that must exist in an ore pulp before flotation can take place: air-mineral contact. When such contact is obtained through the use of the proper collector a froth will form rapidly, and the desired minerals will be found in flocs that consist of mineral particles and minute air bubbles. It is not mutual attraction between mineral particles which causes the formation of such flocs, but adhesion between bubbles and mineral.

Xanthates, by quickly changing the surface of the desired mineral particles to render them water-repellent, produce the essential condition for flotation in a minimum of time. For this reason froth flocculation is often considered characteristic of xanthate collectors. The advantages of rapid flotation may be projected into the design and operation of a mill, as several writers have indicated.

The idea that the *desired* reactions in flotation take place quickly has been offered by Myers and Lewis[1] and by Rose,[2] who point out that undesirable complications may result from dragging out the operation; for example, calcium sulphate deposition, slime activation, slime-coatings, and generation of harmful salts and that the correct application of power in flotation machines would counteract this tendency. In this I concur, but I believe that the choice of collector is of equal importance. Xanthate is a fast worker, consequently its use helps to reduce capital investment, operating costs, and other expenditures which Rose calls the "hidden taxes." Furthermore, froth produced under the favorable conditions described in the preceding paragraph is readily dewatered. In mills where the moisture content of the filtered concentrate is extremely critical because of the methods of transportation, sampling, or feeding at the smelter, the use of a xanthate collector is often conducive to minimum difficulty. Mill tests have demonstrated that the moisture content of flotation concentrate after filtering varied with the type of collector reagent, and that xanthates were found to produce the driest cake.

APPLICATIONS

Generalizations are not looked upon with favor by the hard-bitten millman unless they are supported by actual performance data. For this section of the paper, therefore, statistical data covering the use of flotation collectors (and other relevant matter) in practically all of the sulphide ore-flotation mills of over 100 tons daily throughput in North America were tabulated and analyzed. These will be discussed here under the heads, Differential Lead-zinc Ores (61 mills), Copper (32), Differential Copper-zinc (8), Straight Lead and Zinc (13 and the Tri-State), Gold (22), and Miscellaneous (6). A few omissions occur, resulting from the lack of authentic information, but the data are substantially factual and complete.

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[1] References are at the end of the paper.

TABLE 2—*Flotation Collectors in Differential Lead-zinc Mills*[a]

Size of Mill Tons per Day	All Mills Reported	Mills Using Xanthate as Principal Collector						Mills not Using Xanthate as Principal Collector in Either Circuit
		Both Circuits	Lead Only	Zinc Only	Total Lead	Total Zinc	Total Either	
Number of mills:								
Over 2,000	3	2	1		3	2	3	
1,000 to 2,000	5	4		1	4	5	5	
500 to 1,000	16	8	3	2	11	10	13	3
100 to 500	33	20	1	11	21	31	32	1
Total all groups	57	34	5	14	39	48	53	4
Percentage of all mills		59.6	8.8	24.6	68.4	84.2	93.0	7.0
Tons per day:								
Over 2,000	13,000	11,000	2,000		13,000	11,000	13,000	
1,000 to 2,000	6,550	5,550		1,000	5,550	6,550	6,550	
500 to 1,000	10,580	5,750	1,980	1,250	7,730	7,000	8,980	1,600
100 to 500	7,990	5,000	160	2,480	5,160	7,480	7,640	350
Total all groups	38,120	27,300	4,140	4,730	31,440	32,030	36,170	1,950
Percentage of total tons		71.6	10.9	12.4	82.5	84.0	94.9	5.1

[a] Data from 57 mills in North America with daily rate exceeding 100 tons of flotation feed.

Differential Lead-zinc Mills

Sixty-one mills are included under this heading; 57 of these use the differential process of floating a lead concentrate first and then a zinc product, while the other four float a bulk concentrate of all sulphides, following with differential separation of this concentrate. The main group will be discussed first. Of these 57 mills, 53 use xanthate as the principal collector in one or the other circuit or both and mill 95 pct of the lead-zinc ore now treated by this process in North America (Table 2). In 60 pct of the mills (72 pct of the tonnage) xanthate is the principal collector in *both* circuits; in 68 pct of the mills it is the principal lead collector (82 pct of the tonnage); in 84 pct of the mills it is the principal zinc collector (84 pct of the tonnage). All mills larger than 1000 tons use xanthate as the *principal* collector in one or both circuits.

Of the 39 mills that use xanthate as the principal collector in the lead circuit, 16 employ smaller quantities of a secondary collector. In the zinc circuits, however, only 5 out of 48 mills that use xanthate as the principal collector supplement it with another collector. Generally, xanthate alone appears to do such a complete job in collecting zinc after lead flotation that when a secondary collector is used at all it is more in the billet of a frother than collector. This assumption is supported by the fact that in three mills where xanthate is complemented by a second collector no other frothing agent is used.

It is perhaps advisable to clarify the term "principal collector." In these data we have considered xanthate the primary collector only when it is used in greater quantity than any other collector in the circuit and in most cases more than the total of all others.

In Table 1 is noted the C value of each xanthate. In view of the generally accepted hypothesis that the tendency of a xanthate to be adsorbed on a mineral surface increases with the size of the alkyl group, and the corollary that collecting power increases from ethyl (C_2) to amyl (C_5), it is interesting to note below that most millmen prefer the C_2 xanthate in both lead and zinc circuits.

Of 39 mills using xanthate as principal collector in the *lead* circuit:

Ethyl xanthates are used in	19 mills (48.7 pct)
Mixtures ethyl and amyl or pentasol are used in.....	6 mills (15.4 pct)
Butyl xanthates are used in	6 mills (15.4 pct)
Amyl or pentasol xanthates are used in.............	5 mills (12.8 pct)
Isoprophyl xanthates are used in................	3 mills (7.7 pct)

Of 48 mills using xanthate as primary collector in the *zinc* circuit:

Ethyl xanthates are used in	31 mills (64.6 pct)
Butyl xanthates are used in	6 mills (12.5 pct)
Mixtures ethyl and amyl or pentasol are used in.....	5 mills (10.4 pct)
Amyl or pentasol xanthates are used in.............	5 mills (10.4 pct)
Isopropyl xanthates are used in................	1 mill (2.1 pct)

One striking fact not brought out in this summary concerns the choice between the equally priced and equally stable sodium and potassium ethyl xanthates. The advantage commonly (and not unjustifiably) claimed for the sodium salt is its higher content of xanthogen radical, but despite the economy that should thus result from using the sodium salt, 14 of the 19 users of ethyl xanthate in lead flotation, and 27 out of 31 users in zinc flotation, prefer the potassium salt. It is advisable, therefore, to consider the two alkali salts of any xanthate in the light of distinct reagents; in addition to the difference in xanthogen content, they appear to possess individual characteristics that entitle them to such consideration.

A study of the conditions under which C_4 and C_5 xanthates, or mixtures of C_2 and C_5, are being used in lead circuits indicates that they are preferred when precious metals or copper are present in the crude ore, or when oxidation has so altered the relative responsiveness of galena, sphalerite, and pyrite that a higher alkalinity is necessary for their separation. The preference for C_5 xanthates confirms the findings of Wark and Cox,[3] that amyl xanthate can be used in a higher alkalinity than ethyl, thus leading to a saving by the use of lime in place of soda and also increasing the efficacy of cyanide in the depression of pyrite and sphalerite. Cyanide is an expensive reagent and only the part of it that ionizes is effective in flotation; it can be shown that the number of cyanogen ions obtained from a fixed addition of cyanide increases *threefold* with each increase of 0.5 in the pH of the solution between pH 7.0 and 9.0. Practical millmen take advantage of this fact, for when they use the C_4 and C_5 xanthates in differential lead circuits they are found to prefer lime to sodium carbonate, while with C_2 xanthates sodium carbonate is used in three mills out of four.

In the zinc circuit the preference for the higher-alcohol xanthates is less marked than in the lead, again confirming Wark and Cox,[3] who found amyl xanthate not so effective as ethyl xanthate in separating sphalerite from pyrite except at high alkalinity. Nevertheless, mixtures of C_2 and C_5 xanthates have been found advantageous even at low alkalinities.

Since the successful application of collectors in differential flotation depends upon correct use of the inorganic conditioning agents, a few facts on general practice may be of value. As a pH modifier sodium carbonate is used in 26 lead circuits and 1 zinc circuit, while lime is used in 9 lead circuits and 41 zinc circuits. Neither soda nor lime is used in 22 lead sections or 15 zinc. As regards depressing agents, a summary of the 57 mills indicates:

30 mills use cyanide and zinc sulphate
  (20 with soda and 3 with lime)
11 mills use cyanide, zinc sulphate, and sodium sulphite
  (3 with soda and 2 with lime)
6 mills use cyanide alone
  (2 with soda and 3 with lime)
4 mills use zinc sulphate alone
  (2 with soda)
1 mill uses zinc sulphate and sodium sulphite
5 mills use no depressants (or information is incomplete).

It is significant that in 10 of the 12 mills using sodium sulphite copper occurs in the ore in economic quantity, and in several of these gold is also a contributing factor. There are four general conditions under which cyanide and zinc sulphate may not, by themselves, be effective in depressing sphalerite and pyrite from the lead froth, making necessary the use of sodium sulphite to take the place of all or part of the cyanide:

1. When the water or the ore contains salts of copper or iron the beneficial effect of cyanide may be destroyed or seriously impaired unless a large excess is used.

2. When the ore contains considerable copper in cyanide-soluble form the use of cyanide may cause activation of sphalerite and pyrite rather than depression. Also, some ores through oxidation form ferric sulphate and sulphuric acid during the grinding stage, and these may cause solution of some sulphide copper minerals, thus activating the zinc beyond the power of a reasonable amount of cyanide to counteract it.

3. In some lead-zinc ores chalcopyrite (or tetrahedrite) is particularly sensitive to the depressing action of cyanide, and the use of even a small quantity will result in loss of copper.

4. In lead-zinc ores containing gold the optimum use of cyanide for flotation purposes may cause solution of gold, and consequent loss.

In these cases the substitution of sodium sulphite for cyanide, either wholly or in part, is often found advantageous.

All of the 57 mills use copper sulphate for activation of the zinc mineral. Most of the mills were found to use cresylic acid in the lead circuit and pine oil in the zinc.

Thus far in this section we have considered the differential lead-zinc separation; in addition to the 57 mills using this process there are 4 that use a combination bulk-and-differential process. These 4 mills, with total daily throughput of 1500 tons per day, produce a bulk concentrate of all recoverable sulphides and separate this into two or more products, by the conventional methods already described. Although in theory there are several special conditions (soluble salts, slime interference, pre-activation, low lead content of crude ore, fine intergrowth or association of minerals accompanying coarse liberation of free gangue, and others) that should favor the bulk-differential procedure against the straight differential, in practice the former is limited to ores containing not more than 3 or 4 pct combined lead, copper, and zinc, and a small quantity of pyrite. There is a good reason for this. For substantially complete recovery of a large quantity of desired mineral in the bulk float an excessive amount of collector and frother must be used; the separation step is then found difficult because the relative response of the various minerals has been altered.

In all four mills using the bulk-differential method xanthate is used for the bulk flotation.

A third differential procedure worth mention is that of directly floating a zinc concentrate in the conventional manner, using copper sulphate, and then de-leading the zinc concentrate by floating off a lead product in the presence of cyanide. This is economical and efficient only when the lead content of the crude ore is less than one tenth of the zinc content.

When copper mineral, usually chalcopyrite, occurs in a lead-zinc ore, most of it floats in the lead concentrate; this copper is subsequently recovered as a low-grade matte in the lead blast furnaces, then upgraded and shipped to a copper plant. Because this complication leads to additional treatment charges or reduced payment for copper, the mill may realize a saving by dividing the lead-copper concentrate by flotation. This operation usually is profitable when the copper content of the lead-copper concentrate exceeds one twelfth of the lead assay; for example, 52 pct lead with 4.5 pct copper. There are now six mills in North America that make such a separation, using one of these three procedures:

1. At three mills the lead is depressed by sulphurous acid and caustic starch, then a copper concentrate low in lead is removed after partial neutralization with lime.
2. At two mills the copper is depressed with cyanide while the lead is floated. This method produces clean lead and copper but is more costly than the others, and it may cause losses in dissolved gold if that metal is present.
3. At one mill the lead is depressed with dichromate and the copper is floated. This method is cheap and simple; it effectively removes the copper from the lead, but it does not produce a high-grade copper concentrate. Furthermore, in using this reagent care must be taken to segregate the water resulting from the separation circuit, which otherwise might cause lead losses in the primary section.

The success of all three of these methods hinges upon the collector used in the main circuit, which must be of such nature that it will not interfere with the separation. Xanthates are used in all mills now making the copper separation.

TABLE 3—*Flotation Collectors in Copper Mills*[a]

Size of Mill Tons per Day	All Mills		Mills Using Xanthate as Principal Collector			
					Percentage of Group	
	Number of Mills	Tons per Day	Number of Mills	Tons per Day	Number of Mills	Tons per Day
Over 10,000	8	202,000	3	49,000	37.5	24.3
2,000–10,000	10	35,850	9	31,600	90.0	88.1
500–2,000	7	5,120	7	5,120	100.0	100.0
100–500	7	1,355	5	905	71.4	66.8
Total all groups	32	244,325	24	86,625	75.0	35.5

[a] Data from 32 North American copper mills with daily rate exceeding 100 tons.

COPPER MILLS

The extent to which xanthate is used as the principal collector in North American copper mills is shown in Table 3. Thirty-two mills with total daily throughput of 244,000 tons, or more than 60 million tons of ore per year, are covered by these data.

Of these 32 mills, 24 use xanthate as the principal collector, or 75 pct of the total. In tonnage of ore, xanthates show to less advantage, since only on 35 pct of the copper ore is xanthate used. Analyzing the data in the table we find that 200,000 tons per day, or 83 pct of the total, is milled in only 8 large plants, while the balance of 44,000 is treated in 24 mills under 10,000-ton capacity. In the former group xanthate is used on only 24 pct of the tonnage, and in the latter 89 pct.

The former group is made up largely of the porphyry copper ores.

The porphyry coppers may be roughly divided into two classes, the straight chalcocite ores and the ores containing chalcopyrite. The former ores may be considered more difficult as regards grade of concentrate, because pyrite persists in the froth, partly because it is often coated or stained with secondary copper, and partly because its association with copper mineral is too intimate for physical separation even after minute reduction by regrinding of the primary concentrate. As a result the finished concentrate rarely runs higher than 40 pct copper, although chalcocite itself contains nearly 80 pct copper. Xanthates function well on this type of porphyry ore because potency and speed of collection are the essential properties, and the only purpose of a slower or weaker collector is to reject a low grade, but still valuable, portion of the recoverable mineral when grade of concentrate is unusually important.

In the second class of porphyry ores, the chalcopyritic, the problem is also to separate the copper minerals from pyrite, but in this case there is usually less activation of the pyrite or intimate association of it with copper minerals. Xanthates are used on this type of porphyry ore because they function well in the higher lime alkalinities that often are necessary for maximum selectivity between the pyrite and chalcopyrite.

Of the 24 copper mills in North America using xanthate as the principal collector: 14 (58.3 pct) use ethyl xanthate; 4 (16.7 pct) use amyl or pentasol xanthate; 3 (12.5 pct) use mixture ethyl and amyl or pentasol; 2 (8.3 pct) use butyl xanthate; 1 (4.2 pct) uses mixture amyl and butyl. The sodium ethyl xanthate is preferred to the potassium in 9 of the 14 mills. The C_4 and C_5 xanthates are being used on dump ore or on others where tarnishing or light oxidation has occurred, or where precious metals make the higher cost worth while.

In practically all of the copper mills lime is used as the pH modifier. Cyanide finds application for pyrite depression (and occasionally sphalerite) in five mills, mostly in the cleaning circuit. The use of pine oil as frother is quite general; in a few instances an alcohol frother is used with it.

DIFFERENTIAL COPPER-ZINC MILLS

Eight mills in North America are reported to be making the differential copper-zinc separation. Xanthate is used in either the copper or the zinc circuit at seven mills, and in both circuits at four mills. Lime is more commonly used than sodium carbonate as the pH modifier in the copper section. For deactivating sphalerite and pyrite (or for preventing activation) cyanide is used in six mills; with zinc sulphate in three of these, sodium sulphite in two, and by itself in one.

The separation of chalcopyrite from sphalerite is one of the most difficult, because copper dissolves during grinding and flotation, activating the unwanted minerals. Wark and Cox[3] have indicated that the formation of the copper-bearing film on sphalerite is easier to prevent than to counteract, hence the depressants and pH modifier should be on the ore when it is first pulped with water. They also suggest that this counteraction is best applied before xanthate is added, since the copper-bearing film on sphalerite is more soluble than the dual copper-and-xanthate film.

There is, however, an exception to the last. When an ore contains or develops so much soluble copper that an excessive quantity of cyanide would be needed for its conversion to cuprocyanide, it will occasionally be found more effective and more economical to precipitate some of this copper with xanthate during the grinding period, using up to ½ lb of xanthate per ton of ore, and then to add the cyanide and its companion depressant to the flotation feed. This use of xanthate as a precipitant rather than as a collector is also valuable on some lead-zinc ores.

The separation of the sphalerite from pyrite in the secondary circuit is not usually found difficult with xanthate as the collector. Ethyl xanthate is ideally suited unless a high-lime circuit has been employed for copper removal, in that case butyl or amyl xanthate may be found more efficacious.

LEAD MILLS

Straight lead ore is milled in 10 plants in North America; the largest of these are in eastern Missouri and represent over 90 pct of the total tonnage. Eighty per cent of all the ore is milled in seven mills that use xanthate as the principal collector. Of the xanthates, isopropyl (C_3) is generally preferred for galena flotation at these mills.

ZINC MILLS

The Tri-State district contributes nearly all of the tonnage in this group. The ores contain a small quantity of galena, but only three of the larger mills make a differential separation, these were included with the lead-zinc mills. The others produce only a small amount of lead concentrate on tables or make no effort to produce lead at all. Copper sulphate is used in all of the zinc circuits, and pine oil is the favored frother. With the exception of a few mills in the Tri-State district, all straight zinc operations use xanthate as the principal collector, in nearly all cases C_2 or C_3 xanthate.

GOLD MILLS

Because of the suspension of most of the gold mills in the United States during the war, recent information on operations of this type is not so complete as on the base-metal ores. In the 22 mills covered by this survey the gold most commonly occurs in association with arsenopyrite, some of the gold free or liberated by fine grinding; less frequently the gold is found with pyrite, pyrrhotite, chalcopyrite, or in telluride.

It is a striking fact that all of these mills use xanthate as principal collector. Of the 22 mills with total throughput of 10,000 tons per day: 12 use secondary butyl xanthate; 6 use amyl or pentasol xanthate; 2 use mixtures of ethyl and amyl or pentasol; 2 use ethyl xanthate. This marked preference for the C_4 and C_5 xanthates, despite their higher price, confirms their value as collectors for free gold and for such less responsive sulphides as arsenopyrite, pyrrhotite, tellurides, and tarnished sulphides in general. An interesting detail of gold flotation is the use of silicate or starch to prevent slime interference or to depress iron oxides, mica, or talc; when reagents of this type are found to depress gold-bearing sulphides as well as slime, copper sulphate is used to forestall this tendency.

When a gold ore contains only a small percentage of floatable sulphides, so that their complete recovery would not result in a large tonnage of low-grade concentrate, the metallurgy is relatively simple, for it consists simply of grinding fine enough to liberate the gold and the sulphides from the gangue and then floating them. On the other hand, an ore is occasionally encountered in which the percentage of sulphides is so high that a bulk concentrate would have little value as a shipping product. As an alternative to cyaniding such a concentrate, selective flotation may be found successful if the gold is free, or associated with only a part of the sulphides. I remember such an ore from Oaxaca, Mexico, which contained so much pyrite and arsenopyrite that the gold tenor of the ore could hardly be doubled in the concentrate by bulk flotation. At a pH of 9.4 with sodium carbonate and sodium sulphide, a surprising result was obtained—nearly all of the gold was floated in a small quantity of auriferous pyrite, while nearly all of the barren pyrite, arsenopyrite, and gangue were rejected. It is not uncommon that auriferous pyrite is found more responsive than barren pyrite to xanthate collection, and when it can be selectively floated by careful control of alkalinity, or by the use of such reagents as sodium sulphide, copper sulphate, or dispersants, an otherwise unprofitable operation may be salvaged. Even a simple washing operation before flotation, as on the gold-bearing cerussite ores of northern Mexico, may mark the difference between profit and loss.

Miscellaneous Mills

Under miscellaneous are included two copper-nickel mills, in which C_5 xanthates are used as principal collectors; one tungsten mill, in which a C_4 xanthate is used to float off sulphides before leaching; two mills treating oxidized copper ores, which employ a C_5 xanthate as sole collector; and one acid-leach-float copper mill, in which xanthate is used as the secondary collector. These applications merit an individual, detailed description that is beyond the scope of this paper.

Summary of Applications

Recent data on 170 North American mills shows that 148 of these use xanthate as principal collector in one or more circuits. Table 4 gives a more detailed summary, in percentages. Tonnage percentages are about the same as those shown in Table 4, with only the important exception of the copper mills where xanthate is the chosen collector in 75 pct of the mills but only 35.5 pct of the tonnage.

Table 4—*North American Mills Using Xanthate as Principal Collector in One or More Circuits*

Class of Ore	Total Mills	Total Major Circuits	Xanthate Used as Principal Collector	
			Percentage of Total Mills	Percentage of Total Circuits
Lead-zinc	61	131[a]	93	79
Copper	32	32	75	75
Copper-zinc	8	17[a]	88	71
Lead	10	10	70	70
Zinc	31	31	84	84
Gold	22	22	100	100
Miscellaneous	6	8	83	88
Totals	170	251	85	78

[a] Including iron circuits.

Storage, Handling, Feeding

The xanthates are stable in the sense that they decompose only slowly (2 to 5 pct in a year) when stored in drums in a cool, dry place. It is important, however, to keep xanthate from moisture; if the drums have to be left where they are exposed to rain even temporarily, they should be laid down to keep water from standing on the ends. With sodium ethyl xanthate this precaution is particularly important, for it may take fire spontaneously when damp.

As regards handling, the same ordinary precautions should be taken as with nearly any chemical substance; i.e., if it comes in contact with the skin it should be washed off, it should not be taken internally, and inhalation of its dust or fume (if burning) should be avoided. It is said that an occasional individual is skin-sensitive to xanthate and becomes afflicted with a temporary rash on contact. The recommended treatment is the application of sodium bicarbonate solution and bismuth subnitrate salve, followed by avoidance of further contact for a week or more. Even temporary indisposition from this cause is very infrequent.

Inasmuch as liquids may be fed to the flotation circuit much more easily and accurately than solids, xanthates should be dissolved in water for use. Such solutions are relatively stable, even a 0.2 pct solution shows little hydrolysis after 72 hr. Therefore, the strength of solution can be suited to the operator's convenience. In practice xanthate solutions as weak as 1 pct and as strong as 25 pct are not uncommon. Accuracy is the first requirement of the feeding system and deserves first consideration in determining what concentration is to be employed.

The mechanical reagent feeder is used for xanthate solutions in most mills, but a few of the large mills prefer control by valve, which, though satisfactory with dilute solutions, requires more attention than mechanical feeders. In choosing the feeding system and deciding upon solution strength, the operator should keep in mind that dilute xanthate solutions are not pungent and are, in fact, as inoffensive as a sulphide collector can be.

The solution storage should be large enough for a 24 to 30-hr supply, so that dissolving may all be done on dayshift. A word of caution is added regarding the mixing of these solutions. Dissolved xanthate diffuses slowly, consequently if a tankful is left incompletely mixed it will remain that way for hours. In a recent test 1500 gal of 2 pct xanthate solution, perfunctorily mixed, was found after 24 hr to consist of layers of widely differing concentrations. Filling of the tank *before* xanthate addition will be found helpful; stirring is best performed mechanically or by air, and the minimum period of stirring should be determined in each case by test.

TESTING OF ORES

For an exposition on ore testing refer to Hassialis' Section 19 of Taggart's Handbook.[4] It is a broad subject, and rather than trim it to the scope of this paper I shall confine my comments to a few working rules governing the use of xanthate in ore testing.

A 0.5 pct solution is generally used in the laboratory. Fresh solutions should be made for each day's work; these are often made in tap water and used unfiltered. Preferred practice is to use distilled water. Preliminary tests are run generally with sodium ethyl xanthate until such details as test procedure, fineness of grinding, alkalinity, nature and quantity of inorganic reagents have been established. The factors that deserve most attention in the early phases of the investigation are grind, the degree of alkalinity (or pH) and the quantity of collector. Of these Wark[5] has said that alkalinity is the most important factor in determining the effectiveness of a flotation reagent, while in regard to the collector, Edser once stated that failure in flotation is more likely to occur from using too much rather than too little. Even in a mill of many years successful operation, frequent reexamination of these three factors is profitable.

After a satisfactory procedure has been developed with sodium ethyl xanthate, the other xanthates deserve intensive study, for, as the section on Applications has shown, one will be found superior to the others. Not infrequently, a C_4 or C_5 xanthate produces a good grade of concentrate with excellent recovery but in a froth that is difficult to remove from the machine or to handle in subsequent operations. In this case a mixture with a C_2 xanthate will generally relieve the objectionable condition without sacrifice. However, froth difficulties are not so easily recognizable in the laboratory as in the mill; for this reason a change in xanthate for improvement of froth quality is generally made as a result of large-scale observation.

In regard to the order of addition, xanthate usually is added to the test cell after all other reagents except the frother. When conditioning is to precede flotation, the xanthate may go in either before or after the conditioning period. It may be fed in stages during the roughing period, but ordinarily this is not advantageous in laboratory tests.

The best quantity of xanthate must be determined, naturally, for each ore. It is impossible to forecast or estimate this figure. However, the following approximate averages for different types of sulphide ore may serve as a guide for the preliminary tests:

Type of Ore	Xanthate, Lb per Ton Ore
Lead-zinc, lead circuit	0.10
Lead-zinc, zinc circuit	0.07
Lead-zinc, iron circuit	0.15
Copper	0.04
Copper-zinc, copper circuit	0.08
Copper-zinc, zinc circuit	0.08
Straight lead or zinc	0.05
Gold ores	0.15

At mills treating oxidized copper or oxidized lead, the consumption of xanthate runs as high as 3 lb per ton. Even on sulphide ores, xanthate consumption may run up to a pound per ton when, as described in an earlier paragraph, it is used in the capacity of heavy-metal precipitant rather than as collector. Normally, 0.15 lb per ton is sufficient for collection in a sulphide circuit.

CONCLUSION

We have pointed out that the introduction of xanthate was one of the forces that gave impetus to the progress of flotation shortly after World War I. So effective did selective flotation become in converting millions of tons of waste into ore, and encouraging the development of immense low-grade ore bodies, that several authoritative writers of the early twenties feared that irreparable harm to the mining industry would occur in the form of overproduction. We can now smile at these fears, knowing that our huge metal production during the past seven years was the act for which selective flotation had set the stage 20 years earlier.

Now that we are in a similar period of reconstruction, it is interesting to speculate on the new process developments that must be forthcoming, though as yet unheralded, to mark the second "golden age" of milling progress.

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Effect of "Metso" (Sodium Meta-silicate) on Mill Recoveries of Alta-St. Louis Ores

By Henry P. Ehrlinger,* Member, AIME

(New York Meeting, February 1948)

The Alta and St. Louis mines of the Alta Mines, Inc., produce a somewhat oxidized ore with a talc gangue that presents quite a problem in milling. For several years the mill recoveries were relatively low and there seemed to be no cure for this condition even with increasing the various collecting agents and frothers to two or three times of what they should be. The usual alibi of "oxidized ore" was given as the reason for these results and the matter stood there for some time until these investigations were started.

Flowsheet

As a brief outline of the flowsheet the ore is crushed to minus 1 in. in a Telsmith jaw crusher and then fed to a 64½ Marcy grate mill. The resulting product from the grate mill is passed over a trommel where the coarse particles (plus 7 mesh) are returned to the grinding circuit through the classifier. The under 7-mesh pulp is passed over two Denver mineral jigs where a high grade concentrate is produced, thence over two Wilfley tables where a table concentrate is produced and about 80 pct of the oxide lead recovered. The tailings from these tables are pumped back to the classifier and after classification the 30 pct pulp density product, which overflows the lip, is passed into a series of 10 Denver economy flotation cells where a lead-copper concentrate is produced. Three products are made and shipped to the smelter termed jig, table and float concentrates.

Results with various Reagents

In March of this year work was started on the metallurgical problems of the mill and, as is shown in Table 1, the results obtained were not too efficient.

Table 1—*Mill Results, Alta Mill, March, 1946*

Product	Assays				
	Au	Ag	Pb	Cu	Ox-Pb
Mill heads	0.20	4.8	1.5	0.90	0.8
Jig concentrates	9.20	82.8	24.2	3.6	0.2
Table concentrates	3.30	46.5	25.3	2.1	1.9
Float concentrates	1.01	60.35	11.1	8.3	0.7
Tailing	0.045	1.22	0.7	0.32	0.5
Recoveries, Per Cent					
Jig concentrates	24.6	6.5	18.1	4.2	0.4
Table concentrates	25.9	19.8	21.6	7.6	22.2
Float concentrates	29.4	51.0	18.3	57.0	21.2
Tailing	20.1	22.7	42.0	31.2	56.2

Tests were run with various reagent combinations that were recommended for ores of this nature but there was no appreciable effect on the recoveries. The grades of the flotation concentrates were reduced by the insoluble picked up and retained by these more powerful collectors and frothers.

Results with "Metso"

In past work on ores of this type the use of "Metso" had improved gold and lead recoveries, but no effect on the silver and copper recoveries had been noted. To

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start the tests on a standard basis a large sample of ore was dried and mixed to provide a standard head sample for all tests. The best reagent combination, as shown in Table 2, was used on all tests; the times, pH value, pulp density, and so forth, were held constant with the amounts of "Metso" being the only variable. The reagents, with the exceptions of the "Metso," are shown in Table 2.

TABLE 2—*Reagent Combinations*

Reagent	Amount per Ton Ore Milled	Introduced At
Aerofloat 31 and Sharples 124	0.050	Ball mill
Z_8 xanthate	0.350	Classifier pool
Aerofloat 404	0.350	Ball mill
Cresylic acid	0.020	5th cell

PERTINENT DATA

Water pH	7.6
Grinding time, min	20
MOG	12.6 pct plus 48 mesh
Pulp density of grind	55 pct solids
Pulp density flotation cell	30–24 pct solids
pH flotation pulp	8.0
Flotation time min	30

The amounts of "Metso" used for each test and the assays of the products for each test are shown in Table 3 and the recoveries for each test with "Metso" are given in Table 4.

TABLE 4—*Recoveries with "Metso" by Tests*

PER CENT

Test Number	Au	Ag	Pb	Cu	Zn	Ox-Pb
1	89.6	88.9	78.5	76.4	21.3	29.2
2	91.1	94.2	78.1	80.6	57.3	52.5
3	91.3	92.8	77.8	80.4	56.6	76.7
4	94.4	94.2	83.5	85.3	72.6	88.4
5	96.3	94.2	89.2	90.2	72.3	88.3
6	96.3	93.8	89.0	90.0	71.6	88.0

These recoveries were sufficient to justify a trial of "Metso" in the mill. The rate of feed at one pound per ton seemed the best. These results indicated that with 1 lb of "Metso" at $0.05 per pound gold recovery was increased from 89.6 to 96.3 pct, a gain of $0.52 per ton; silver recovery was increased from 88.9 to 94.2 pct, a gain of $0.31 per ton (at 70¢ silver); lead recovery was increased from 78.5 to 89.2 pct, a gain of $0.15 per ton; copper recovery was increased from 76.4 to 90.2 pct, a gain of $0.37 per ton.

The "Metso" addition was started in the mill in September 1946, at the rate of 1 lb per ton of ore milled. The results obtained in the mill over a 45-day period, milling over 5000 tons of ore, are shown in Table 5.

TABLE 3—*Amounts of "Metso" Used and Assays of Each Test*

Test Number	"Metso" Lb per Ton	Product	Au	Ag	Pb	Cu	Zn	Ox-Pb
All		Heads	0.230	6.1	0.80	0.90	0.30	0.35
1	0.00	Concentrate	0.950	35.1	4.60	4.85	1.15	0.90
		Tailing	0.030	0.8	0.30	0.25	0.25	0.28
2	0.25	Concentrate	1.050	37.3	5.10	5.65	1.15	1.10
		Tailing	0.025	0.6	0.25	0.20	0.15	0.20
3	0.50	Concentrate	1.150	46.5	5.20	6.20	1.25	1.40
		Tailing	0.025	0.5	0.25	0.20	0.15	0.10
4	0.75	Concentrate	1.250	50.5	5.80	6.35	1.25	1.60
		Tailing	0.015	0.4	0.20	0.15	0.10	0.05
5	1.00	Concentrate	1.300	60.3	6.20	7.50	1.30	1.75
		Tailing	0.010	0.4	0.10	0.10	0.10	0.05
6	2.00	Concentrate	1.440	65.4	7.10	8.20	1.40	1.80
		Tailing	0.010	0.4	0.10	0.10	0.10	0.05

TABLE 5—*Mill Results, Alta Mill, September–October*, 1946

Product	Au	Ag	Pb	Cu
COMPOSITE ASSAYS				
Mill heads	0.153	4.25	1.21	0.70
Jig concentrates	13.530	24.15	76.91	0.81
Table concentrates	9.020	42.20	54.50	1.28
Float concentrates	0.980	65.60	12.54	11.95
Tailings	0.008	0.50	0.15	0.05
RECOVERIES, PER CENT				
Jig concentrates	28.9	1.1	17.6	0.5
Table concentrates	37.5	8.5	38.7	1.5
Float concentrates	28.1	81.9	35.5	85.5
Tailings	5.5	8.5	8.2	12.5

The results over this period clearly indicated that the use of this reagent on the ore was fully justified. In November 1946, the silver tailing had dropped to 0.40 oz per ton, thus bringing mill results up to those obtained in the laboratory work.

SUMMARY

The particles of gold and galena at the Alta and St. Louis mines were badly coated with slime after grinding and the addition of metasilicate had the effect of defloculating the slime in the pulp and cleaning the surfaces of the mineral particles so they were more ammenable to flotation. Sodium silicate was used in one test but the results were not improved to any extent.

Effects of Oxidation of Sulphide Minerals on Their Flotation Properties

By Enid C. Plante* and K. L. Sutherland*

(New York Meeting, March 1948)

Practical metallurgists are unanimous in stating that oxidation of mined sulphide ore adversely affects separation of the constituent minerals under standard conditions in a mill. Frequently, the need for different operating conditions in different plants is attributed either to the extent of oxidation of the mineral surfaces or to soluble oxidation products. The investigations reported in this paper study the effect of oxidation on the primary condition necessary for flotation, namely that of adhesion of an air bubble to the mineral.

Oxidation Products from Sulphide Minerals

Weinig and Carpenter[1] state that the relative rates of oxidation of some sulphides are:

$$FeAsS > FeS_2 > CuFeS_2 > ZnS > PbS > Cu_2S \text{ tetrahedrite.}$$

They also show that oxidation is rapid at first, that sulphides of older geological formations are more stable than those of recent date and that mixtures oxidize more rapidly than any one mineral alone.

The more rapid oxidation of mixtures is borne out by results of Gottschalk and Buehler[2] who found that PbS, ZnS and CuS oxidized 8 to 20 times faster when pyrite was present. Only the end products of oxidation, sulphate and the metal ion, were identified in these tests.

Gaudin[3] states that pyrite is difficult to oxidize in the laboratory. Winchell,[4] using pyrite containing 99.98 pct FeS_2, allowed water to percolate through a bed of this ground material. At the end of three months only a trace of Fe^{+++} and $SO_4^{=}$ were found. Even after 10 months the quantity of Fe^{+++} was only 7.8 mg and $SO_4^{=}$ 25.5 mg per liter of percolating solution. Carmichael[5] confirmed Winchell's results on the slow oxidation of pyrite.

Usually the oxidation of sulphide surfaces has been conducted in neutral or acid solutions, the acidity being caused by the oxidation products. Eliseev and Nagirnyak[6] found that treatment of pure pyrite with lime gave only thiosulphate, but considered it possible that thiosulphate was obtained not by oxidation of the pyrite itself, but by oxidation of calcium disulphide. Mitrofanov[7] suggested that thiosulphate was formed by the following process: oxidation of sulphide to give free sulphur (this frequently happens with marcasite), followed by oxidation of the sulphur to sulphur dioxide and reaction of the sulphur dioxide with sulphur to give thiosulphate. Ferric and ferrous hydroxides, or similar compounds, were shown to be present.

The Research Staff of the United Verde Copper Co.[8] treated ore from the United Verde mine with lime. Thiosulphate, sulphide and polysulphides were stated to be

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[1] References are at the end of the paper.

formed but no specific tests for these ions were reported. Pyrite was postulated to react to form CaS, CaS_x, CaS_2O_3 and $Fe(OH)_3$. This treatment with lime decreased the flotability of the pyrite. The presence of soluble sulphide has not been claimed by any other investigators.

Mitrofanov and Benenson[7] showed that in grinding chalcopyrite the important factors with respect to the production of soluble oxidation products were pH value, temperature, time of grinding, ratio of liquid to solid and time of aeration. A similar study was made for bornite and covellite.[9] The general conclusions were:

1. As the pH value increased the amount of sulphur-containing acids appearing in solution increased greatly. The consumption of oxygen was greater in alkaline than in neutral or acid solutions and was greater with lime as alkali than with caustic soda.
2. Increase in temperature decreased oxidation both with regard to the amount of oxidizable anions formed and the amount of sulphur (as sulphate and sulphur-containing ions) in solution.
3. Increasing the pulp density decreased the percentage of surface oxidized, because of insufficient oxygen.
4. The amount of oxidation increased with time of aeration either during grinding, pre-agitation (conditioning), or flotation.

Before attempting to assess the influence of different oxidation products of sulphur on flotation some knowledge of the stability of the polythionates is necessary.

The data determined by Kurtenacker et al.[10,11] and those presented by Newton-Friend[12] show that sulphide, sulphite, sulphate and dithionate are stable at pH values between 3 and 12. The following decompose in acid solution: thiosulphate (below a pH of 5), tetrathionate (below 8.9), trithionate (below 5.6). Pentathionate and hexathionate are stable only at pH values of less than 2.5 and 0.3, respectively. Trithionate is stable at pH values above 8.9 and tetrathionate between 8.9 and 11.4. Since most sulphide flotation is conducted in the range of pH values between 6.5 and 11 the stable ions which may be present are $SO_4^{=}$, $S^{=}(SH^-)$, $SO_3^{=}$, $S_2O_6^{=}$, $S_3O_6^{=}$, $S_4O_6^{=}$, $S_2O_3^{=}$. It is also noteworthy that at pH values between 7 and 8 polythionates react with cyanide[13], e.g.

$$S_4O_6^{=} + CN^- + 2OH^- \rightarrow S_2O_3^{=} + SO_4^{=} + CNS^- + H_2O$$

so that in practice cyanide may be consumed when added.

EXPERIMENTAL METHODS

Determination of Soluble Oxidation Products

For the tests the purest minerals obtainable were ground and suspended in water of a known pH value through which air (or oxygen) was bubbled for a definite time at a constant temperature. Usually the pulp density and the surface area per unit weight of the mineral were smaller than those common in flotation.

Analytical methods and the minimum concentration of ion which could be detected are given in Table 1.

TABLE 1—*Summary of Analytical Methods*

Ion to Be Determined	Method	Sensitivity, Mg Per l	Page Number in Snell[14]
Fe^{+++}	Thiocyanate	0.01	283–298
Fe^{++}	Thiocyanate after oxidation with bromine or nitric acid	0.01	283–298
Pb^{++}	Dithizone	0.05	202–204
Cu^{++}	Diethyldithiocarbamate	0.01	164–166
Ag^+	Turbidimetric as chloride	0.005	409–410
Zn^{++}	Dithizone	0.1	350–352
$SO_4^{=}$	As barium sulphate by the Tyndall Effect (a Zeiss Tyndallometer was used)	0.1	613
Ions containing sulphur	Oxidized by bromine and determined as sulphate.	0.1	

The determination of thiosulphate, sulphite, trithionate and tetrathionate, in the presence of one another was attempted, but no satisfactory method was developed.

Qualitative tests for these anions have been described in the literature.[12,15]

In reporting results the total amount of thiosulphate, sulphite, trithionate and tetrathionate is recorded as the amount of sulphate ion which is produced after oxidation with bromine. These anions are hereafter referred to as the thio-salts.

The pH values of the solutions were determined using indicators.

Captive Bubble and Cylinder Flotation Tests

For all tests, mineral as pure as possible was taken from massive specimens. The mineral was then polished or crushed under grease-free conditions. For captive bubble tests specimens of mineral were boiled with a solution of sodium hydroxide (to remove grease) acid (to remove NaOH) and water. They were ground under water on glass plates with graded aluminum oxide polishing powders, then polished under water on velvet pads with aluminium oxide.[16–18] The specimens were handled with glass tongs and kept under water in glass beakers. Rubber gloves were worn during polishing and all glassware was cleaned with chromic acid before use. The velvet polishing pads were first extracted by benzene and then by a solution of sodium hydroxide to free them from grease. After polishing, the specimens were wiped under water with linen pads which had also been extracted with benzene and sodium hydroxide. The surface was tested for cleanliness by bringing air bubbles into contact with various parts of it for 30 sec and only if the bubbles showed no signs of "clinging" or "catching" was the specimen considered clean for use.

Mineral for cylinder flotation tests was prepared by boiling it with a solution of sodium hydroxide, acid and water and then crushing it under water with a porcelain pestle and mortar, previously cleaned with chromic acid. It was then washed into a clean, glass-stoppered 50 ml graduated cylinder and shaken with water and terpineol. Only if no mineral floated was it used for tests. In this paper "unoxidized" refers to mineral which was used within two hours of being polished or ground. "Oxidized" mineral was allowed to stand for two days in aerated distilled water at a known pH value—unless otherwise stated this was the natural pH value of the water.

The test solutions were prepared by adding reagents in the following order: acid or alkali, cyanide, water, other reagents, xanthate, water to standard volume. The pH value was then adjusted to the required value using the glass electrode or indicator.

The mineral was well washed with the test solution and conditioned in the solution for 30 min at 35° ± 1°C. It was then tested for contact or flotation. The polished specimens were wiped with linen pads before being tested for contact to remove any precipitate from the surface. Terpineol (purified)—10 or 20 mg per liter—was used as a frother in the cylinder flotation tests. The pH value of the solutions was measured again after the conditioning period.

From the results of cylinder and contact tests curves have been drawn separating solution compositions in which flotation (contact) occurs from those in which it does not. Fig 26 to 31 show the experimental points obtained and are typical examples of how all curves were constructed. Where two curves appear on the one diagram it is not possible to show experimental points without confusion.

Potassium ethyl xanthate was prepared in the laboratory, purified by crystallization from alcohol[17] and checked by titration against iodine.[16] Fresh xanthate solutions were prepared every second day. The concentration of the cyanide solution was checked regularly by titration against silver nitrate (Liebig's method). The dithionate, trithionate and tetrathionate were prepared in the laboratory by the methods described in the literature.[12,19,20] The purity (at least 99 pct) of the samples was confirmed by a determination of the sulphur and alkali metal content. Other inorganic reagents were of Analar grade.

RESULTS OF OXIDATION TESTS

Qualitative Results

Table 2 shows the products which were formed when various sulphides were suspended either in pure water (pH value, 6) or in an alkaline solution (pH value between 10 and 11) through which air was bubbled for two days at room temperature (15° to 18°C).

TABLE 2—*Products formed by Oxidation of Aqueous Suspension of Sulphide Minerals*

Mineral	Products in Suspension, Originally Neutral (pH = 6)	Products from Alkaline Suspension (pH = 10 to 11)
Galena	*Present:* Cu^{++}, Fe^{++}, Fe^{+++}, SO_4^- *Absent:* Pb^{++}, polythionate	*Present:* H^+, Ag^+, Pb^{++}, SO_4^-, thio-salts *Absent:* Fe^{++}, Fe^{+++}, Cu^{++}
Pyrite	*Present:* Fe^{++}, H^+, SO_4^- *Absent:* Fe^{+++}, polythionate	*Present:* H^+, SO_4^-, $S_3O_6^-$, $S_4O_6^-$, $S_2O_3^-$, SO_3^- *Absent:* Fe^{+++}, Fe^{++}, Cu^{++}, S^-, $S_5O_6^-$, $S_2O_6^-$
Chalcopyrite	*Present:* H^+, Cu^{++}, Fe^{++}, Fe^{+++}, SO_4^- *Absent:* Polythionate	*Present:* H^+, SO_4^-, $S_2O_3^-$, $S_4O_6^-$ *Absent:* Cu^{++}, Fe^{++}, Fe^{+++}, SO_3^-
Sphalerite	*Present:* SO_4^-, thio-salts *Absent:* Zn^{++}, Fe^{++}, Fe^{+++}, Cu^{++}	*Present:* H^+, Zn^{++}, SO_4^-, SO_3^-, $S_4O_6^-$, $S_2O_3^-$

Polythionates are only formed in alkaline solution. With the exception of the solutions containing suspended galena, heavy metal ions are not formed in alkaline solutions. The relative proportions of the soluble products vary with time, which suggests either that the mineral surface becomes coated with a thick nonoxidizable film, e.g., heavy metal hydroxide or basic sulphate which hinders further attack, and/or that the soluble thio-salts react with one another to produce secondary products which may also influence the oxidation of the mineral.

The problem was not studied sufficiently critically to distinguish between these possibilities since the object of the investigation was to identify the soluble oxidation products and obtain the order of their concentration in solution. They could then be tested at these concentrations for their effect on flotation of the minerals.

Quantitative Results

Table 3 shows the amount of oxidation products formed by oxidation of aqueous suspensions of sulphide minerals under the following conditions: temperature—15° to

TABLE 3—*Quantity of Soluble Oxidation Products Formed by Oxidation of Aqueous Suspensions of Sulphide Minerals.*

SUSPENDED IN DISTILLED WATER

(pH VALUE = 6)

Oxidation Product	Concentration (mg per liter) of product formed from			
	Pyrite	Chalcopyrite	Sphalerite	Galena
SO_4^-........	30		2	1
Thio-salts....	Nil	Nil	24	Nil
Cu^{++}........		4.2	Nil	Nil
Fe^{+++}.......	0.01	1	Nil	Nil
Fe^{++}........	15	24	Nil	Nil
Pb^{++}, Zn^{++}..			Nil	Nil
pH (final)....	3.4	2.6	6.0	6.0

SUSPENDED IN WATER TO WHICH SODIUM HYDROXIDE WAS ADDED TO GIVE AN INITIAL pH VALUE OF 10.1.

Product	Concentration (mg per liter) of product formed from			
	Pyrite	Chalcopyrite	Sphalerite	Galena
SO_4^-........	40	20	30	30
Thio-salts....	60	90	70	30
SO_3^-........	Nil	Nil		
$S_2O_3^-$........	Nil	5		
Fe^{++}, Fe^{+++}, Cu^{++}, Zn^{++}	Nil	Nil	Nil	Nil
Ag^+, Pb^{++}...				3.6, 1.1 (respectively)
pH (final)....	5.3	7.2	9.6	5.5

18°C; preparation of mineral—ground in a mortar to −36 mesh and deslimed; pulp density—one part of mineral to 5 parts of water; and time of aeration—five days. Either distilled water or water to which sodium hydroxide had been added was used. Air, freed from carbon dioxide, was bubbled fairly rapidly through the suspension. Although the minerals were ground without much attention to their final size, it will be shown later that mineral area (i.e., size) is relatively unimportant.

The behavior in alkaline solution is almost independent of the mineral, the greater portion of the oxidation products being derived from the sulphur.

Variables Affecting Amount of Soluble Oxidation Product

1. *Solid/Liquid Ratio*[a]—For ratios between 0.05 and 0.25 the amount of oxidation products were constant.

2. *pH Value of Pulp*—Table 3 shows the effect of an increase in pH value. A further increase to a pH value of 12 using pyrite, increased the amount of thio-salts (estimated as sulphate) from 60 to 600 mg per liter.

3. *Temperature*—One comparison made with galena at 15° and 100°C showed only a small increase in amount of soluble oxidation products, e.g., thio-salts increased from 130 to 160 mg per liter and Pb^{++} from 0.6 to 3.5 (pH value of test 10.7).

4. *Size (area of mineral)*—Identical amounts of oxidation product were produced from +60 and −60 mesh mineral.

5. *Effect of Adsorbed Film*—An adsorbed film of cetyl trimethylammonium ion on galena did not alter the rate of oxidation.

6. *Effect of Solutions Buffered by Carbonate*—Larger amounts of thio-salts were formed when solutions were buffered because decrease in pH value was prevented.

7. *Time of Aeration*—No accurate measurements were made on the rate of formation of oxidation products with time. The amount increased and the ratio of thio-salts to sulphate increased with time, indicating that the rate of oxidation of mineral to thio-salts is greater than the rate of oxidation of these to sulphate.

8. *Rate of Aeration and Oxygen Content of Admitted Gases*—Provided aeration was sufficiently brisk to keep the pulp almost saturated with air, variations in rate of aeration did not cause any difference in the amount of oxidation products formed. This is consistent with the normal course of actions of this type in which the rate of diffusion of the soluble reactant—oxygen—to the surface of the solid governs the rate of reaction. A four-fold increase in concentration of oxygen supplied to the pulp (by using pure oxygen instead of air), produced an increased rate of oxidation of pyrite, but the increase was not proportional to the increased oxygen supply.

The effect of variation is solid/liquid ratio, temperature and mineral size also supports the hypothesis that the rate-determining step is the actual concentration of oxygen at the mineral surface, this concentration being determined by the rate of diffusion of the oxygen through a thin liquid film not greatly influenced by agitation. At high temperatures the decreased solubility of air is apparently offset by the more rapid diffusion of the oxygen to the mineral surface. Variation in mineral size will not affect the amount of oxidation products if the solid/liquid ratio is sufficiently large, since most of the oxygen will be removed from the immediate neighborhood of the mineral.

Concurrent with the study of the above variables, tests were conducted to determine the effect of the oxidation products on the flotation of sulphide minerals investigated in this paper. These tests showed that the oxidation products in the amounts formed were unimportant, and the factors listed above were therefore not studied further.

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[a] The actual data whereon these conclusions are based are available from the authors.

Effect of Oxidation Products on Air—Mineral Contact

Effect of Cations

The effects of silver, lead, zinc,[21] and copper,[22] have already been studied with xanthate as collector. Fig 1 compares the effect of sodium cyanide as a depressant for the various sulphides in the presence of 167 mg $FeSO_4 \cdot 7H_2O$ per liter (stoichiometrically equivalent to 150 mg $CuSO_4 \cdot 5H_2O$ per liter). Fig 2 records similar data for ferric ammonium sulphate. The ferric salt is particularly effective in increasing the range of pH values in which a separation of chalcopyrite from pyrite is possible.

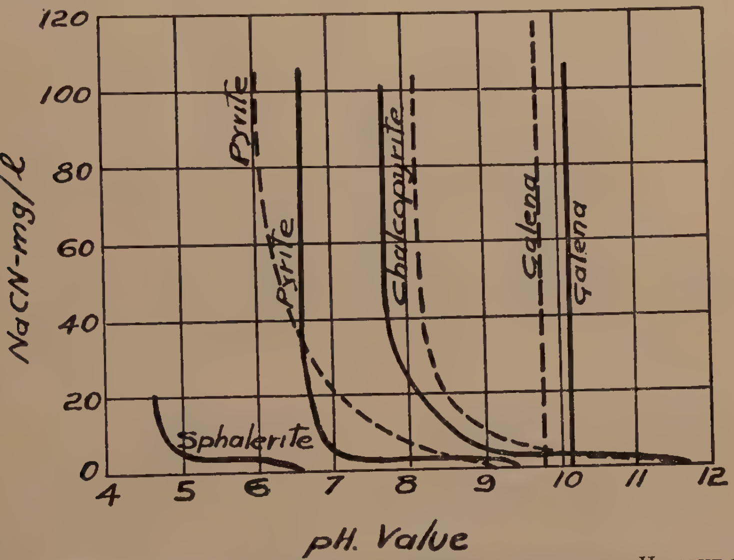

Fig 1—Relationship between concentration of sodium cyanide and pH value necessary to prevent contact at mineral surfaces.

KEtX = 25 mg/l. $FeSO_4 \cdot 7H_2O$ = 167 mg/l. Temperature = 35°C.

Air-mineral adhesion is possible only in solutions the compositions of which are given by points to the left of or below the curve for that mineral. Broken curves show results in the absence of ferrous sulphate.[23] Sphalerite does not float.

The flotation of sphalerite (Fig 1) could not be caused by copper activation since: (1) Small amounts of cyanide would de-activate it; (2) when the ferrous salt was dissolved in water and boiled with pure iron and sulphuric acid, the copper content was reduced to less than 10^{-3} mg per liter for a 1 gm $FeSO_4 \cdot 7H_2O$ per liter solution. This solution still activated sphalerite in the presence of xanthate.

Activation caused by ferrous salts falls within a different category from that by lead, copper or silver *if*, the behavior of these reagents is judged by their ability to form insoluble sulphides and xanthates. Thus ferrous ethyl xanthate is soluble to the extent of about 1 pct in water whereas ferrous sulphide is the most soluble of the sulphides of zinc, iron, lead and copper.[24]

Ferrous salts also differ from the more usual activators in that pretreatment will not activate a mineral such as sphalerite or galena.

If ferrous sulphate is added in conjunction with copper sulphate (Fig 3), the conditions of separation of sphalerite from pyrite are improved. Carbonate was not

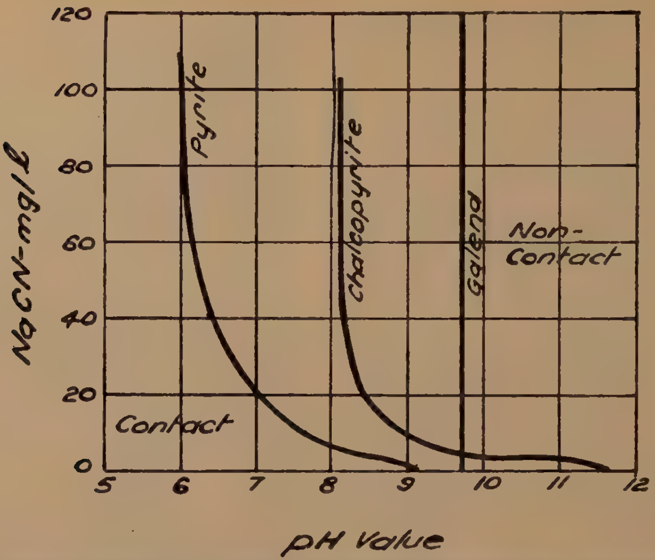

Fig 2—Relationship between concentration of sodium cyanide and pH value necessary to prevent contact at mineral surfaces.
KEtX = 25 mg/l. $Fe_2(NH_4)_2(SO_4)_4 \cdot 24H_2O$ = 290 mg/l. Temperature = 35°C.

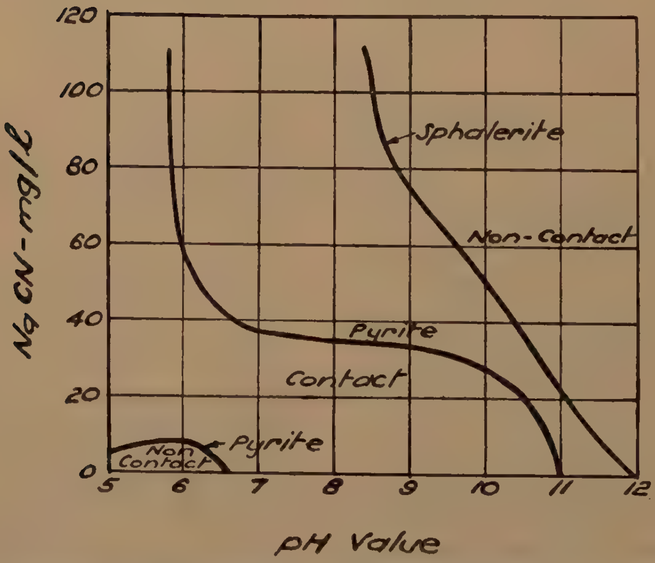

Fig 3—Relationship between concentration of sodium cyanide and pH value necessary to prevent contact at mineral surfaces.
KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 167 mg/l. Temperature = 35°C.

added in these tests and the xanthate concentration was low. Fig 4 illustrates the behavior of sphalerite and pyrite with a lower concentration of ferrous sulphate. Under these conditions separation of these minerals would not be possible.

The general effect of ions such as copper, lead and silver in the absence and presence of cyanide is understood.[16,21,22]

Results with zinc, ferrous and ferric ions are more difficult to interpret. Galena is activated by ferrous salts, unaffected by ferric salts; chalcopyrite activated by ferrous and ferric salts, pyrite depressed by ferrous and unaffected by ferric salts. No evidence could be found for the existence of iron xanthate complexes which could be used to explain these results. Solutions in which both ferrous and copper salts are present or are added, are particularly valuable for the separation of sphalerite from pyrite, but no simple explanation of their behavior could be adduced.

Effect of Anions

The anions thiosulphate, dithionate, trithionate, tetrathionate and sulphate had no effect on the flotation of chalcopyrite, galena, pyrite or sphalerite whether unoxidized or oxidized with ethyl xanthate as collector. Thiosulphate and trithionate caused no depression of sphalerite in the presence of copper, cyanide and xanthate.

Sulphite had no effect on the flotation of chalcopyrite, galena or sphalerite but depressed pyrite (Fig 5). It has been reported[a] that with ethyl xanthate as collector and at a pH value between 7 and 8 sulphite is an effective depressant for galena and pyrite but allows chalcopyrite to be floated. Since sulphite ion is not a depressant for galena in solutions containing only ethyl xanthate, the effect of added copper was investigated. In the presence of a solution containing 25 mg $CuSO_4 \cdot 5H_2O$ per liter, 150 mg Na_2CO_3 per liter, and 5 mg potassium ethyl xanthate per liter at pH values from 6.7 to 8.5 neither pyrite nor galena floated, whereas chalcopyrite floated normally. At a pH value of 6.7 galena began to float. A half-hour conditioning period was necessary for efficient flotation of chalcopyrite.

To ensure that all anion compounds which might be derived from the oxidation of a sulphur mineral do not affect flotation, the filtrates from pyrite and other minerals oxidized in alkaline solutions were used to make up collector solutions instead of distilled water. These solutions behaved as though prepared with distilled water.

EFFECT OF OXIDATION UPON THE MINERAL SURFACE AND ON FLOTATION

Gaudin[25] found that "pure" pyrite is very floatable[a] when fresh and poorly so when oxidized. Eliseev and Nagirnyak[6] confirm this by grinding pure pyrite in air or oxygen for different periods of time. The oxygen-treated mineral floated less readily than that ground in air. Tests by Malinovski[26] had shown that long aeration after grinding increased the flotability of pyrite but Eliseev and Nagirnyak showed that if the time of aeration was decreased and the mineral floated with nitrogen or oxygen instead of air, then with nitrogen eight times more pyrite floated than with oxygen. They concluded, therefore, that the intermediate oxidation products on a mineral surface were not detrimental to flotation but that the final product at the pyrite surface, namely sulphate, adversely affected flotation.

Malinovski's results are suspect however because his tests were made with a pyritic ore containing copper minerals from which copper salts were certainly obtained as the sphalerite also became floatable as the oxidation proceeded. Iron and zinc salts might also be expected in solution but they are less marked in their effect than copper.[16]

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[a] Private communication from Lake George Mines Ltd., Captain's Flat, N.S.W. Australia.

[a] "Flotability" is used in the sense of recovery of a given mineral under constant conditions of flotation.

In a second paper[27] Malinovski claimed that the better flotation of his pyrite was due to removal of ferric hydroxide and basic sulphates by nascent sulphuric acid. This is similar to a claim by Gaudin[28] that if oxidized pyrite is treated with sodium sulphide flotation is improved. Even better results were reported if sulphuric acid was added to the mineral. It was considered that these treatments either removed the "oxide" layer or converted it to the original state.

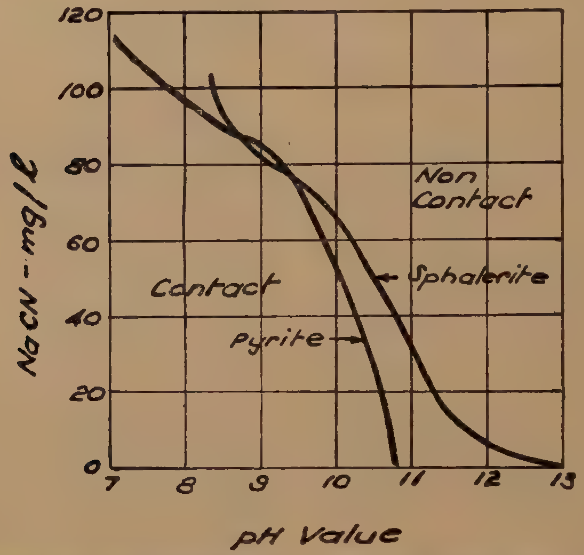

FIG 4—RELATIONSHIP BETWEEN CONCENTRATION OF SODIUM CYANIDE AND pH VALUE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 16.7 mg/l. Temperature = 35°C.

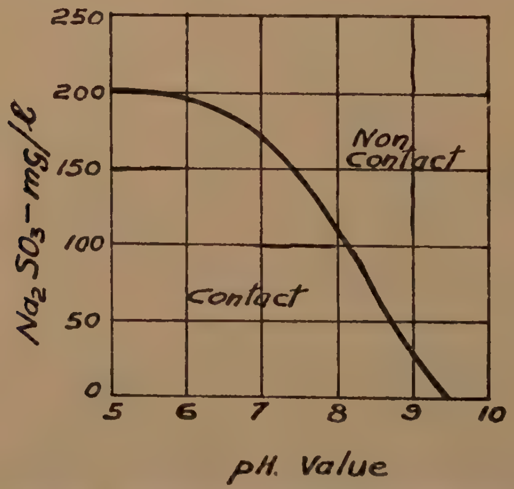

FIG 5—RELATIONSHIP BETWEEN CONCENTRATION OF SODIUM SULPHITE AND pH VALUE NECESSARY TO PREVENT CONTACT AT A PYRITE SURFACE.

KEtX = 25 mg/l. Temperature = 35°C.

The Research Staff of the United Verde Copper Co.[8] suggests that decreased flotability of pyrite shown in their tests on a pyritic copper ore was caused by films of solid materials such as ferric hydroxide. Similarly McLachlan[29] found that oxidation of ore helped to float the chalcopyrite and reject pyrite.

The flotation of selected chalcopyrite was studied by Mitrofanov and Benenson,[7] who found that the important factors influencing the amounts of soluble oxidation products were pH value, temperature, time of grinding, pulp density and time of aeration. The flotation of chalcopyrite, however, was unaffected by oxidation for the reagent conditions used.

Siedler[30] has shown that the floatability of galena oxidized by exposure to air and water decreased enormously as oxidation was continued. Ravitz and Porter,[31] Ravitz,[32] Herd and Ure,[33] and Wark[16] have considered in detail the effect of an "oxide" film at the surface of galena. It was contended particularly by Ravitz and Porter that such a film is essential for adsorption of xanthate.

Knoll[34] investigated the oxidation products formed at the surface of galena, and found that addition of xanthate at high concentrations liberated OH^- and/or $CO_3^=$, $SO_4^=$, and reducing ions of the type $S_mO_n^=$ $(n/m < 4)$. No sulphide ion was liberated but thiosulphate was shown to be present among the displaced ions. The results of tests by this author showed clearly that a galena surface is not entirely PbS.

Thus, with the exception of the results obtained by Malinovski, it would appear that flotation of a sulphide is hindered by excessive oxidation of its surface. Malinovski's results may be due to copper activation.

EXTENT OF OXIDATION OF MINERAL SURFACE

The factors which may influence the *extent* of oxidation of a mineral surface are:

1. Temperature of the solution.
2. Oxidizing agent used, e.g., MnO_4^-, MnO_2, O_2, I_2.
3. Accumulation of oxidation products at the mineral surface which may hinder continued oxidation.
4. pH value of the solution.
5. Time of aeration.

The factors which may influence the *rate* of oxidation are:

1. Temperature of the solution.
2. Rate of supply of oxidizing agent.
3. Rate of accumulation of oxidation products.
4. pH value of the solution.

The rate of oxidation of the surface is not considered in this paper, the conditions being chosen so that properties of the surface did not change significantly on further contact with the oxidizing agent. This was accomplished by leaving the mineral in aerated water until, when tested in a solution containing 25 mg potassium ethyl xanthate per liter, the critical pH value required to prevent flotation (air-mineral contact) did not alter as the time of oxidation was further increased. This period varied between 2 and 4 days depending upon the mineral, pH value of the aerated water, temperature and rate of aeration. Thus for pyrite the critical pH values required to prevent contact in the presence of 25 mg KEtX/l. were:

Freshly polished specimen	10.0
Oxidized for 1 day	10.5
Oxidized for 2 days	10.8
Oxidized for 4 days	10.8
Oxidized for 8 days	10.8

Specimens of freshly cleaved pyrite, pyrite reduced cathodically during polishing or pyrite pretreated in a solution containing sodium sulphide before testing all gave the same critical pH value as the freshly polished specimen. Hence either the specimens prepared by polishing are unoxidized or the initial oxidation of the surface is complete within 2 or 3 min. Specimens of pyrite which were polished and then stored in oxygen-free water for three days showed a critical pH value of 10.0, the same as that for a freshly prepared specimen.

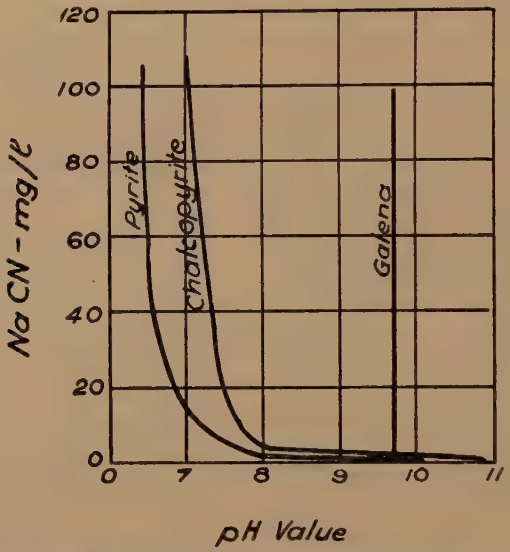

Fig 6—Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at mineral surfaces.
KEtX = 25 mg/l. Temperature = 35°C.
In this and the following figures, unless otherwise indicated, —— = unoxidized mineral - - - = mineral oxidized in water of initial pH value of 6. (Fig 6 is taken from Wark and Cox.[23])

Unless otherwise specified all mineral was oxidized at room temperature (15° to 20°C) for a period of two days, this being a sufficient time in freely aerated water. The distilled water used in the oxidation had a pH value of about 6 which fell to 4 or 5 as the oxidation proceeded. In other tests the minerals were oxidized in an alkaline solution of pH between 10.5 and 11.0, which resulted in a different subsequent behavior. To facilitate distinction between curves for oxidized and unoxidized mineral, curves for oxidized mineral are shown by broken lines in the figures.

Effect of Oxidation on the Response to Xanthate Using Hydroxyl and Cyanide as Depressants

Fig 6 and 7 show that galena, pyrite and chalcopyrite were all more tolerant to additions of cyanide and alkali after oxidation of the mineral surface in distilled aerated water. Fig 8 shows similar results for specimens oxidized in an alkaline solution. After this treatment the chalcopyrite was more tolerant to cyanide whereas pyrite was less tolerant. From these results it is obvious that a mixture of unoxidized and oxidized chalcopyrite cannot be separated from unoxidized and oxidized pyrite using cyanide and alkali as depressants with 25 mg potassium ethyl xanthate per liter as collector.

Mineral oxidized at 10° or 35°C showed the same behavior as did that oxidized at room temperature.

Effect of Oxidation on Response to Xanthate Using Copper as Activator and Alkali and Cyanide as Depressants

The specimens were oxidized in distilled water and the response of the mineral to xanthate in the presence of copper, cyanide, hydroxyl but in the absence of carbonate[22] was determined. Fig 9 and 10 compare the behavior of the minerals when 25 mg

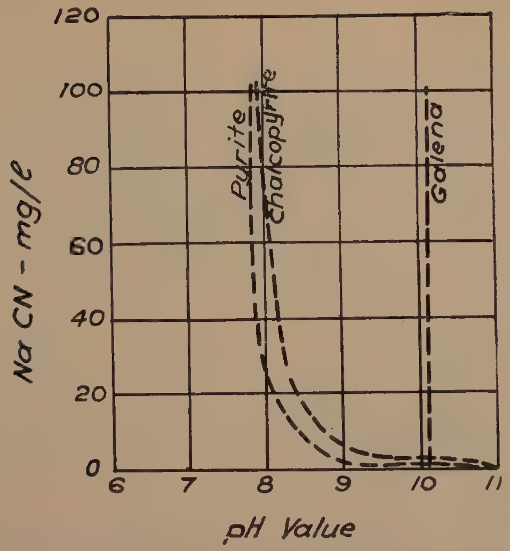

FIG 7—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.
KEtX = 25 mg/l. Temperature = 35°C.

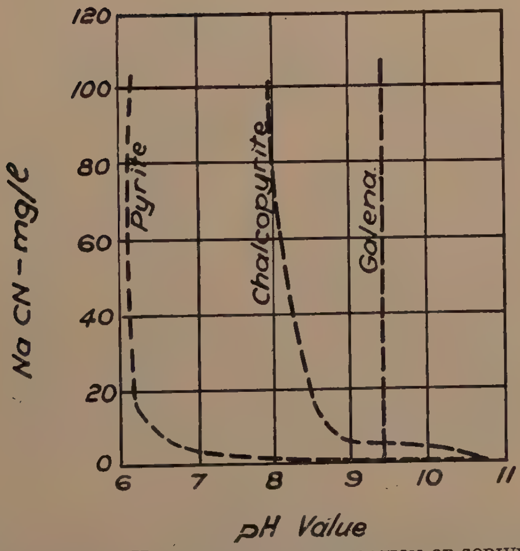

FIG 8—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.
KEtX = 25 mg/l. Temperature = 35°C.
Mineral oxidized in water at pH 10.7.

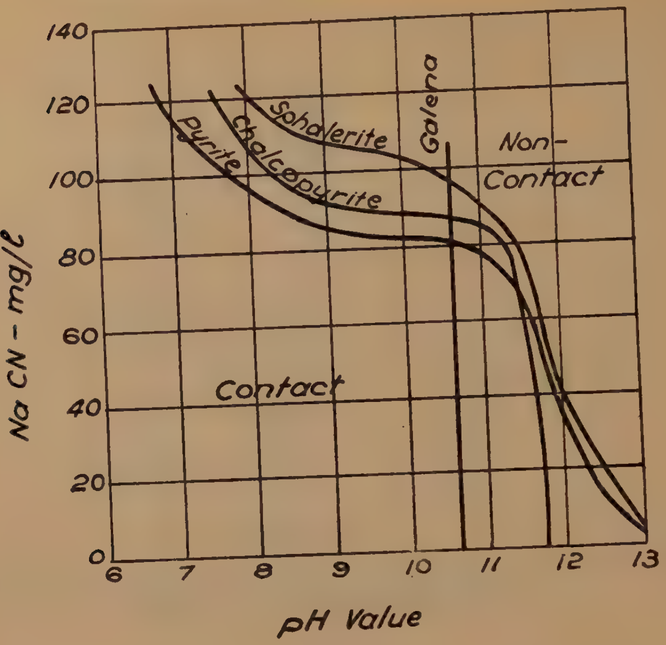

FIG 9

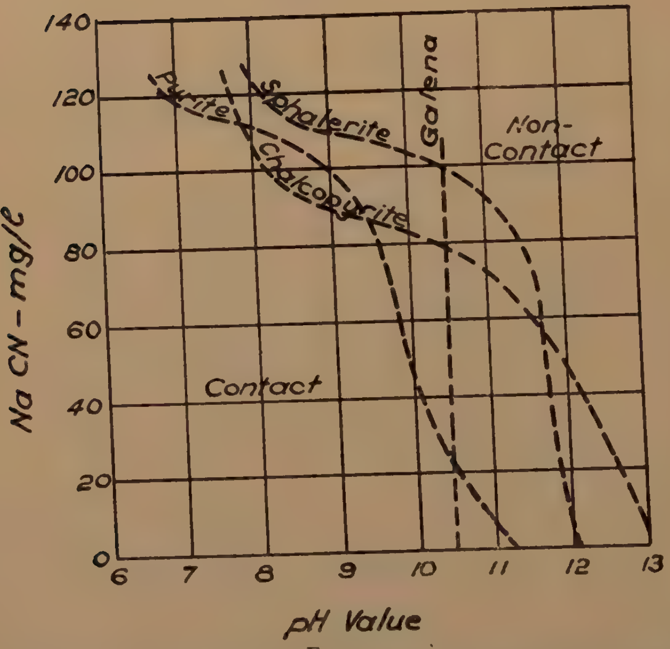

FIG 10

FIG 9 AND 10—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 25 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = nil. Temperature = 35°C.
Fig 9 = unoxidized. Fig 10 = oxidized.

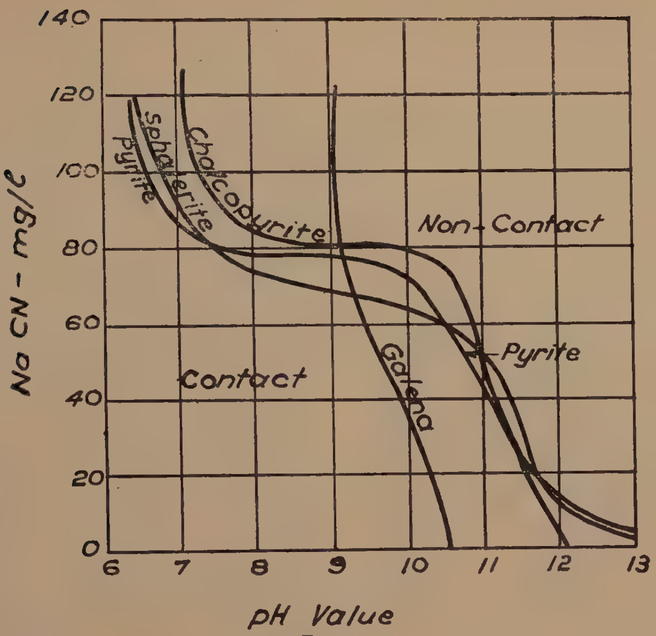

FIG 11

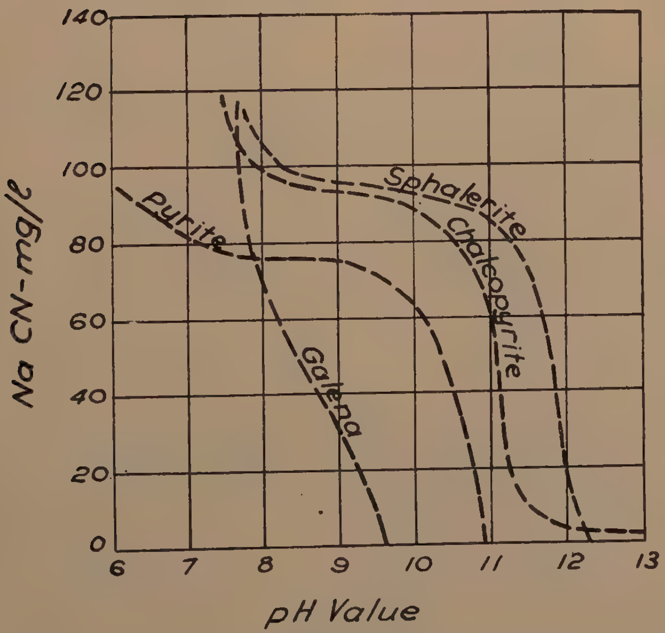

FIG 12

FIG 11 AND 12—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = nil. Temperature = 35°C.
Fig 11 = unoxidized. Fig 12 = oxidized.

potassium ethyl xanthate per liter was used as collector. Fig 11, 12, 13, 14 and 15 show the results when 5 mg of collector per liter was used, carbonate being present for the tests recorded in Fig 14 and 15.

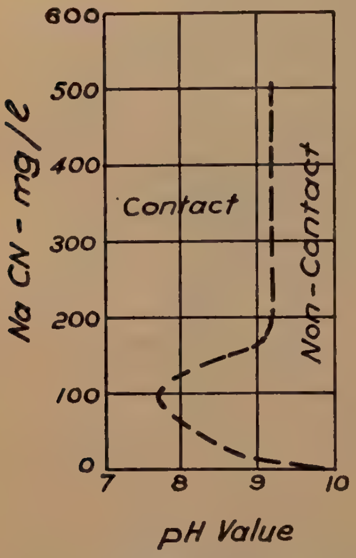

FIG 13—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT A GALENA SURFACE.
KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

The critical pH value for galena in the presence of 5 mg potassium ethyl xanthate per liter and in the absence of copper is 9.2. Usually, if copper sulphate is present then as soon as the cyanide concentration is sufficient to give the soluble complex $Cu(CN)_2^-$ the film of copper is removed from the galena and thereafter cyanide has no depressant effect; flotation (contact) will therefore cease at the same pH value as in a test where neither cyanide nor copper salts are added.[35] For oxidized galena with 5 mg potassium ethyl xanthate per liter and 150 mg $CuSO_4 \cdot 5H_2O$ per liter this independence to cyanide is not shown until the concentration exceeds 180 mg per liter (Fig 13). To demonstrate whether the galena possessed a copper-bearing film, the specimen was washed in distilled water and the critical pH value determined in a 25 mg potassium ethyl xanthate per liter solution free from copper and cyanide. When specimens were copper-activated the critical pH exceeded 9.7, whereas when unactivated it is 9.7. If the addition of NaCN exceeds 50 mg per liter the mineral is not activated. Since cyanide is not a depressant for unactivated oxidized galena, it seems certain that the $Cu(CN)_2^-$ is a depressant for the oxidized galena. As the concentration of cyanide is increased this ion is converted into a second complex $Cu(CN)_4^{\equiv}$ which is not a depressant for galena.

Usually sphalerite is separated from pyrite in solutions containing copper sulphate. The reagent conditions used for determining the curves reported in Fig 14 and 15 show that it would be feasible to separate sphalerite from pyrite when these minerals are present both in the unoxidized and oxidized state (a cyanide concentration of 40 to 60 mg per liter in the pH range 7.5 to 8.5 would be satisfactory.) In the absence of the carbonate, separation of sphalerite from pyrite is not possible (Fig. 11 and 12).

The "island" of noncontact and the form of the pyrite curves (Fig 14 and 15) is caused essentially by xanthate deficiency (the copper precipitates the xanthate) and not by lack of activation by copper. Oxidation does not alter the mineral markedly so that the conclusions presented regarding island formation may still be regarded as valid.[22]

EFFECT OF OXIDATION ON RESPONSE TO XANTHATE IN PRESENCE OF FERROUS SALT AND ALKALI AND CYANIDE AS DEPRESSANTS

Ferrous salts activate galena, chalcopyrite and pyrite (Fig 1). Oxidation of these minerals enhances the activating effect of the ferrous salt (Fig 16 and 1). Separation of galena, chalcopyrite and pyrite from each other would be very easy if all the

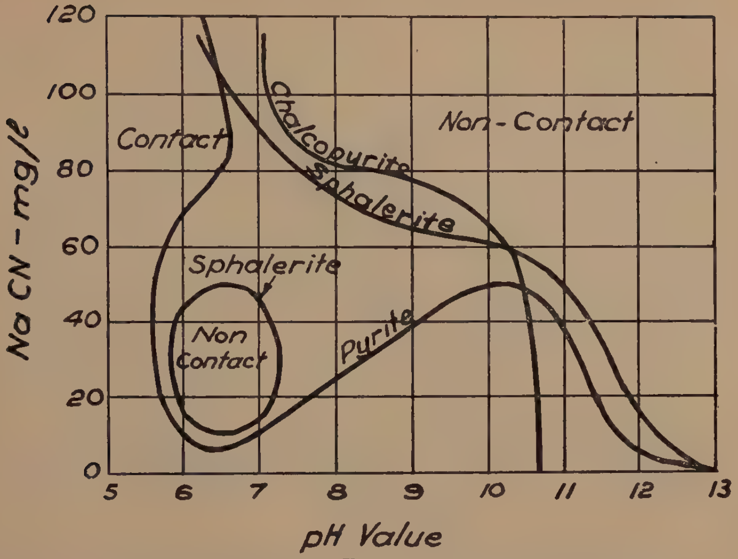

FIG 14

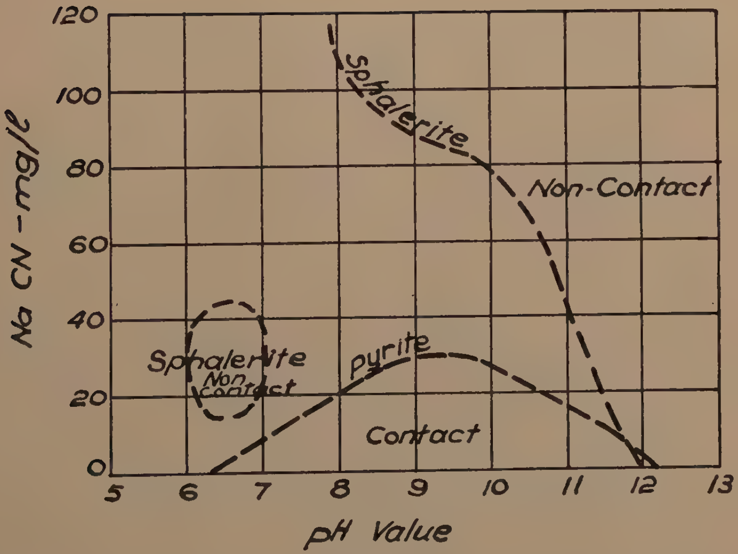

FIG 15

FIG 14 AND 15—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = 150 mg/l. Temperature = 35°C. Fig 14 = unoxidized. Fig 15 = oxidized. (Fig 14. After Wark and Sutherland.[22])

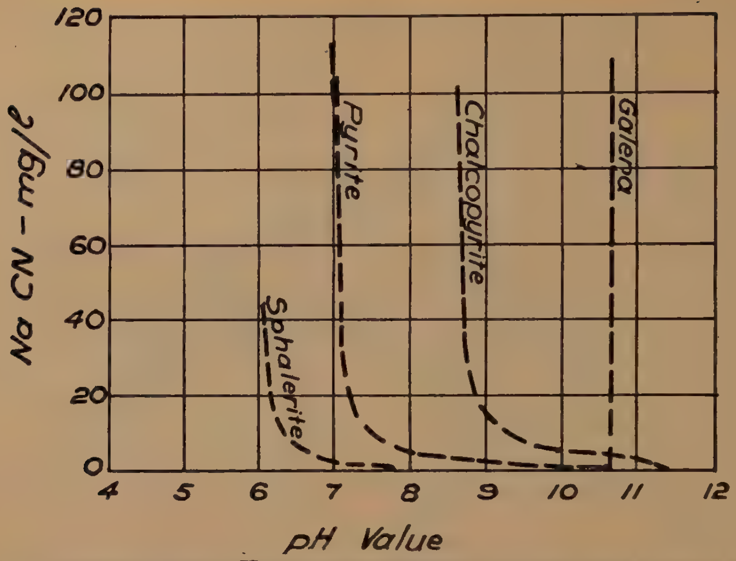

FIG 16—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT OXIDIZED MINERAL SURFACES.
KEtX = 25 mg/l. $FeSO_4 \cdot 7H_2O$ = 167 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

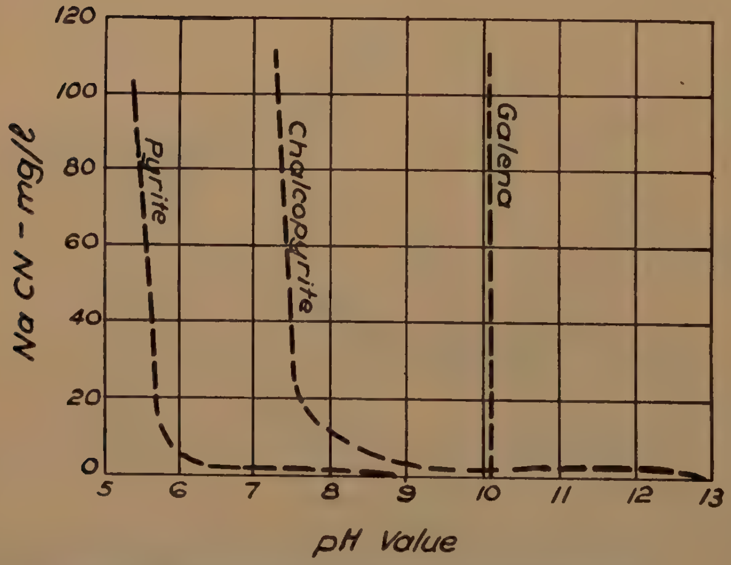

FIG 17—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT OXIDIZED MINERAL SURFACES.
KEtX = 25 mg/l. $Fe_2(SO_4)_3(NH_4)_2SO_4 \cdot 24H_2O$ = 290 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

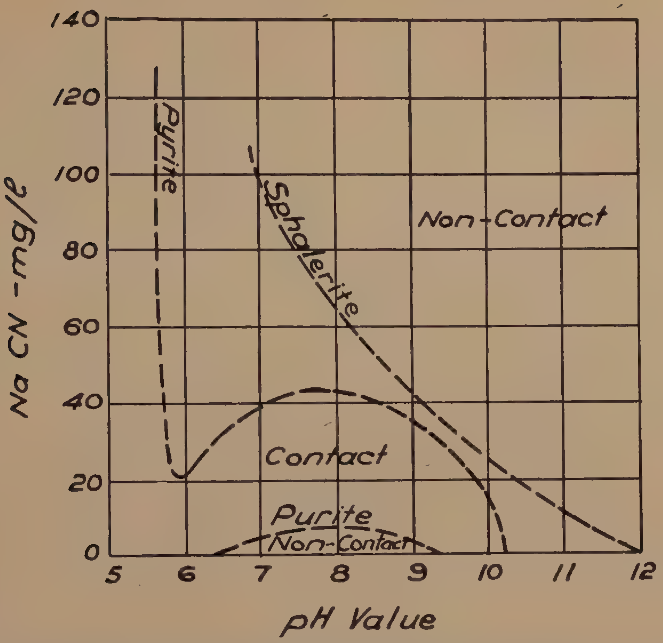

FIG 18—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT OXIDIZED MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 167 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

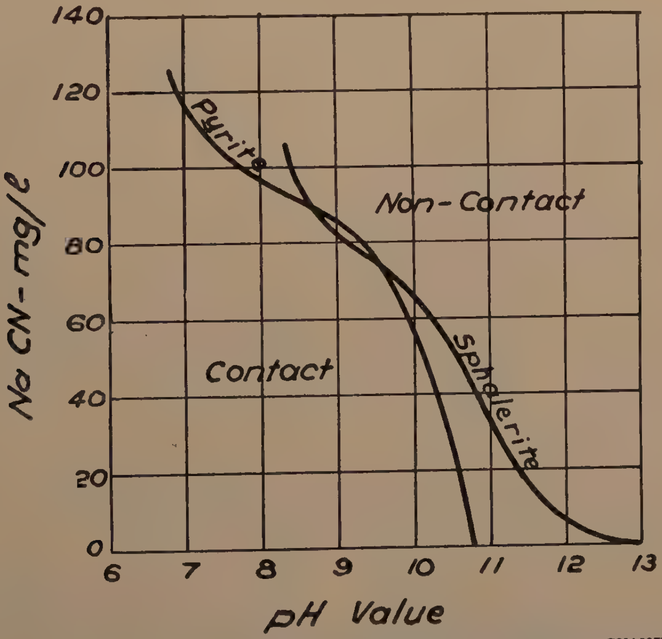

FIG 19—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT OXIDIZED MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 16.7 mg/l. Temperature = 35°C.

mineral could be obtained either in the unoxidized or oxidized state. If there is a mixture of both oxidized and unoxidized minerals, the separation would still be possible but difficult.

EFFECT OF OXIDATION ON RESPONSE TO XANTHATE IN PRESENCE OF FERRIC SALT AS ACTIVATOR AND ALKALI AND CYANIDE AS DEPRESSANTS

The effect of ferric salts on unoxidized mineral is shown in Fig 2. Oxidation of the minerals increases the depression of pyrite and depresses chalcopyrite (Fig 17), but separation of mixtures containing both oxidized and unoxidized minerals might present difficulties. Unoxidized or oxidized galena and chalcopyrite could be readily separated from pyrite.

EFFECT OF OXIDATION ON RESPONSE TO XANTHATE IN PRESENCE OF COPPER AND FERROUS SALTS AS ACTIVATORS AND ALKALI AND CYANIDE AS DEPRESSANTS

Fig 18 and 3 show that the addition of 167 mg of ferrous sulphate per liter to a system containing copper, no carbonate and low xanthate (compare Fig 11 and 12) causes marked depression of pyrite whether oxidized or unoxidized. Pyrite can even be depressed in the absence of cyanide.

Sphalerite is less affected by the addition of ferrous salt although oxidation diminishes its tolerance to cyanide and separation of sphalerite from pyrite, may be possible at high cyanide concentrations. If the concentration of ferrous salt is reduced to one-tenth (Fig 19), the behavior of sphalerite is very similar to its behavior at the high concentration of ferrous salt but pyrite is not depressed at all by the smaller addition. A separation of these minerals is then not possible. This demonstrates how a variation in the concentration of salts not normally considered important may affect flotation.

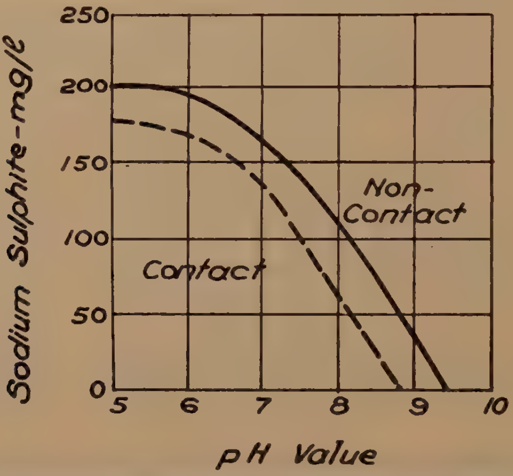

FIG 20—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM SULPHITE NECESSARY TO PREVENT CONTACT AT AN OXIDIZED PYRITE SURFACE.
KEtX = 5 mg/l. Temperature = 35°C.

EFFECT OF OXIDATION PRODUCTS ON THE FLOTATION OF OXIDIZED MINERALS

Chalcopyrite, pyrite and galena were crushed separately and placed in N/100 alkali solution for 5 days while air free from carbon dioxide was passed through the suspension. The filtered oxidation liquids were used instead of distilled water to make up solutions for cylinder flotation tests with these oxidized minerals. The results were the same as those with solutions made with distilled water. This confirmed tests with dithionate, trithionate, tetrathionate and thiosulphate which were not depressants for pyrite, chalcopyrite or galena. Fig 20 shows that sulphite depresses pyrite but when the minerals are oxidized it is not formed in sufficiently large amounts to cause depression.

POSSIBLE METHODS OF OVERCOMING DIFFERENCES BETWEEN OXIDIZED AND UNOXIDIZED ORE

Inhibition of Oxidation

It was shown earlier (page 169) that if mineral was stored in oxygen-free water there was no change in the surface for a period of several days. Therefore, if a reducing agent is added to the pulp during grinding it might be possible to prevent oxidation. Ions which have no detrimental effect on subsequent flotation should be considered (e.g., $SO_3^{=}$). Fundamentally the method is difficult because the moist ore is oxidized rapidly during mining, storage and crushing as well as during grinding. No attempts were made, therefore, to apply the method.

Completion of Oxidation

If the sulphide minerals in an ore have their surfaces uniformly oxidized then the problem of separation by flotation is simplified because it concerns oxidized mineral only. Unfortunately the rate of oxidation with air is slow. Use of oxidizing agents during grinding, e.g., permanganate, dichromate or manganese dioxide would probably give more uniform oxidation, although there is no guarantee that the extent of oxidation of the surface would correspond to that produced by oxidation with air or oxygen. Residual products from the oxidizing agent would also have to be considered.

In a private communication from the American Cyanamid Co., a method of separation of pyrite, pyrrhotite and arsenopyrite by use of the oxidizing agent, potassium permanganate, was described. It is not certain that the whole action is caused by the oxidation of the mineral by permanganate, since any permanganate which is not consumed by the mineral will oxidize the xanthate to dixanthogen. The object of the tests was to produce a pyrite and an arsenopyrite concentrate with a maximum rejection of pyrrhotite. The percentage recovery of arsenopyrite and pyrite as separate concentrates was 90 pct, the concentration ratios being respectively 84:1 and 196:1. In a similar test omitting the treatment with permanganate it was not possible to separate the arsenopyrite from the pyrite. In this example use was made of the more marked difference in behavior of the oxidized minerals.

Removal of Oxidized Surface

The third method which might lead to a uniform surface is by treatment of the oxidized and unoxidized mineral with sulphide ions (or a combination of reagents which either contain or can supply sulphide) to restore the surface to its unoxidized condition. The method adopted was to prepare oxidized and unoxidized mineral, immerse both in a solution of sodium sulphide (20 mg) for 10 min, wash well with distilled water, and determine the behavior of the minerals by captive bubble tests in the usual way. Washing was necessary since sulphide in solution is an excellent depressant with ethyl xanthate as collector.[36]

Pretreatment of a freshly polished surface with a 20 mg $Na_2S{\cdot}9H_2O$ per liter solution produced no change in the response of unoxidized galena, pyrite or chalcopyrite to a solution of 25 mg potassium ethyl xanthate per liter. When oxidized minerals were similarly pretreated, the response of galena and pyrite became identical with that of the freshly prepared surface, whereas chalcopyrite responded as though the surface remained partly oxidized, i.e., the behavior was intermediate between that of the oxidized and the unoxidized mineral. Although the above method is effective, therefore, for the depression of pyrite when making a combined copper and lead concentrate, it suffers from the disadvantage of requiring adequate washing of mineral to remove sulphide. Since ferrous iron is effective in activating chalcopyrite and galena but not pyrite when the mineral is unoxidized (Fig 1), it may be feasible to precipitate most of the preactivating sulphide as ferrous sulphide in alkaline solution. Fig 21 and 22 illustrate the results when 25 mg potassium ethyl xanthate per liter was used as collector in a solution containing sodium sulphide and ferrous sulphate. The amount of ferrous sulphate is stoichiometrically in excess of the sulphide. The mineral specimens were pretreated for 10 min in a solution of sodium sulphide, washed and put into the test solution. The reagent combination causes oxidized or unoxidized pyrite and chalcopyrite to behave almost identically, and separation is possible under the same conditions as those for fresh mineral. Higher additions of ferrous salt should improve the conditions for separation and improve the behavior of galena which is depressed in the presence of the sulphide. The extreme sensitivity of galena to the depressant effect of the hydrosulphide ion was shown by Wark and Cox.[36]

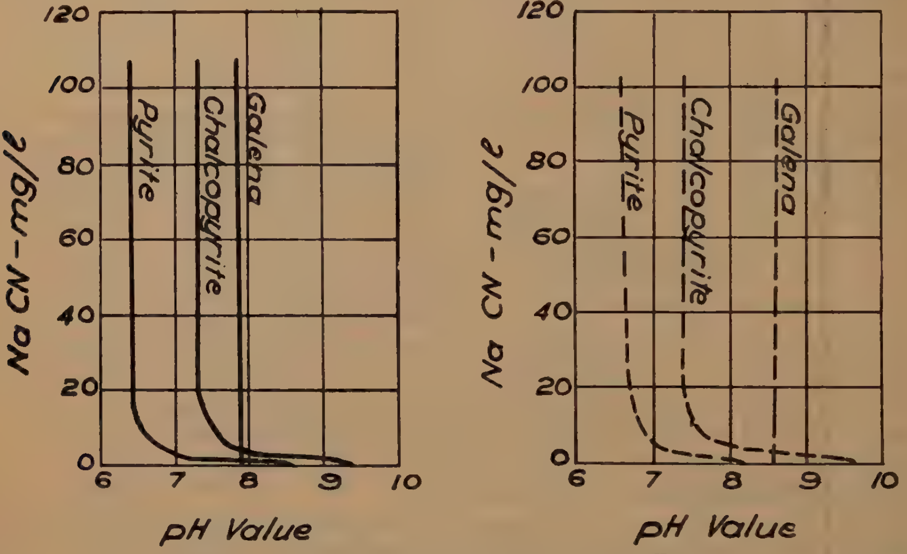

Fig 21 Fig 22

Fig 21 and 22—Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at mineral surfaces.

KEtX = 25 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

Fig 21 = unoxidized. Fig 22 = oxidized.

In the absence of carbonate, separation of partly oxidized pyrite in the presence of copper sulphate (150 mg per liter) was not possible whether the xanthate concentration was high or low (Fig 9 to 12). Pretreatment of the oxidized mineral with sulphide produced a response to xanthate (Fig 23) which theoretically allows a separation. However, it is doubtful if this sepa-

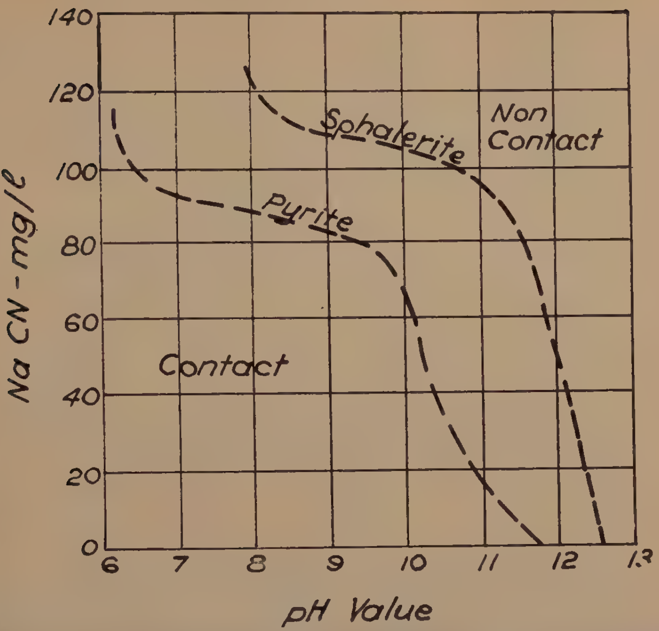

FIG 23—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 25 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Oxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

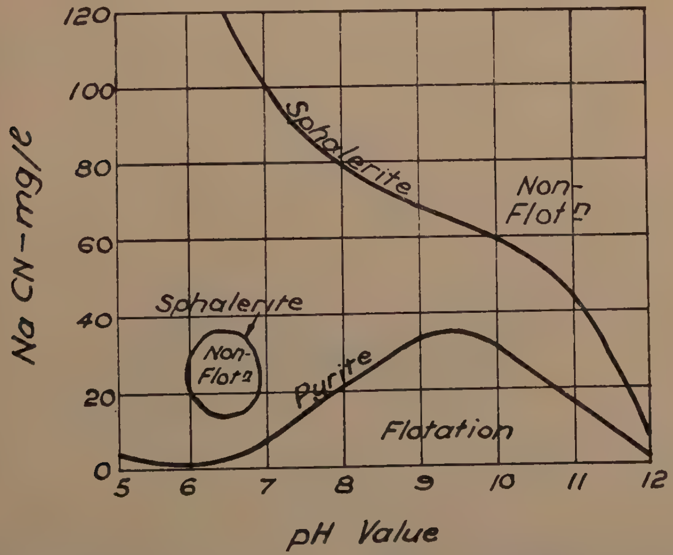

FIG 24—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = 150 mg/l. Temperature = 35°C.

Unoxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

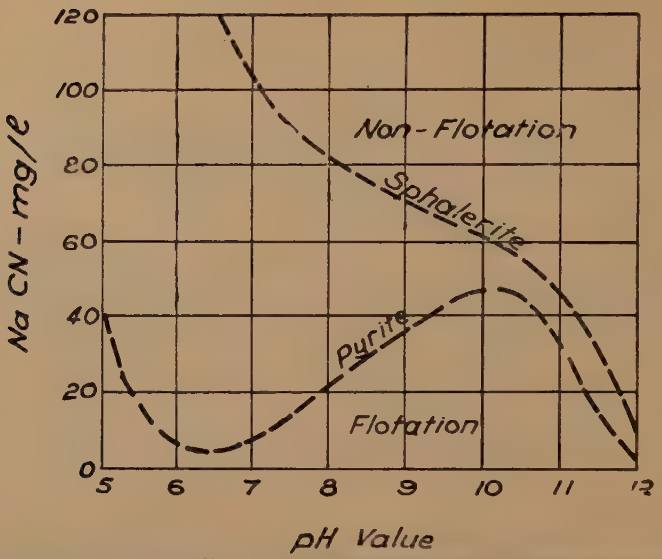

FIG 25—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT MINERAL SURFACES.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. Na_2CO_3 = 150 mg/l. Temperature = 35°C.

Oxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

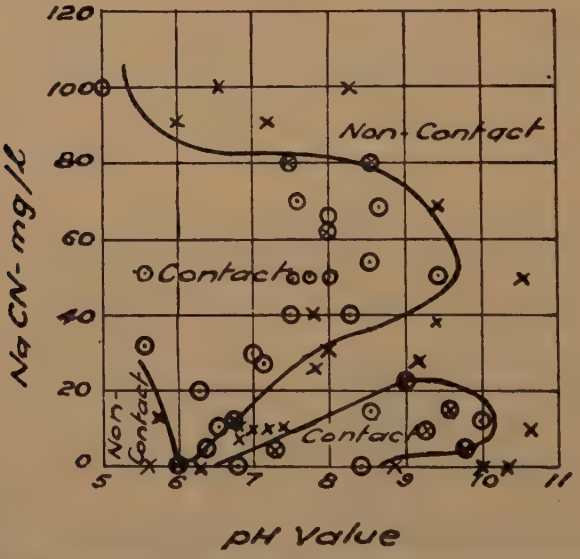

FIG 26—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT A PYRITE SURFACE.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Unoxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

O = air-bubble contact.
⊗ = "catching" of an air-bubble.
X = no air bubble contact.

ration would be practicable as both pH value and cyanide concentration must be high.

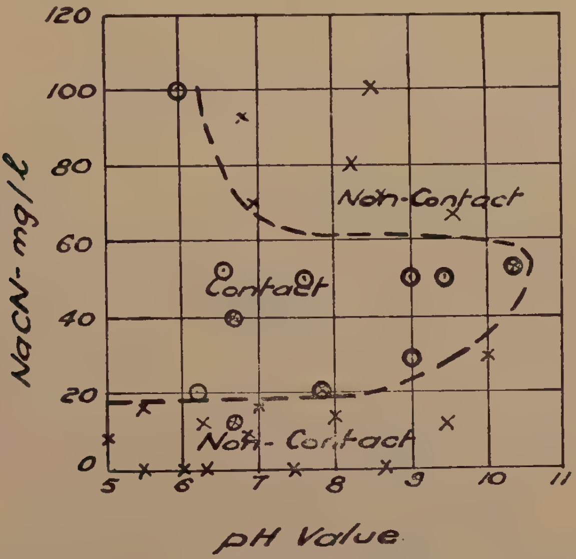

FIG 27—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT A PYRITE SURFACE.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Oxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

O = air-bubble contact.
⊗ = "catching" of an air-bubble.
X = no air-bubble contact.

If the concentration of xanthate was lowered and carbonate added (Fig 14 and 15), separation of sphalerite (oxidized and unoxidized) from pyrite (oxidized and unoxidized) was possible without pretreatment with sulphide, but such pretreatment considerably increases the range of conditions for separation (Fig 24 and 25). Whatever method is adopted the cyanide concentration is more critical than the pH value.

Fig 26 to 31 show the results when ferrous sulphate and sodium sulphide were used in conjunction with copper sulphate, low xanthate and no carbonate for minerals pretreated with a sodium sulphide solution. The ferrous and copper sulphates were not effective in removing the added sulphide, and it would be impossible, for example, to separate partly oxidized sphalerite from partly oxidized pyrite. The curves are more complex than usual and possibly the solutions are not in equilibrium since the minerals were given only the usual 30-min conditioning period. Longer conditioning periods appeared undesirable from a practical angle. The sequence of the curves for each mineral seems logical and the nonflotation or noncontact when cyanide is absent (e.g., chalcopyrite and spalerite) points to the effective depressant action of sulphide. Sulphide depresses galena, sphalerite and chalcopyrite strongly whereas pyrite can tolerate relatively high concentrations of sulphide. For pH values between 7.5 and 10.5 (Fig 28) chalcopyrite was depressed by the sulphide (absence of cyanide), whereas pyrite can be floated (see island of flotation, Fig 26). The sulphide becomes less effective at pH values outside this range because only the ion SH^- is effective as a depressant.[86]

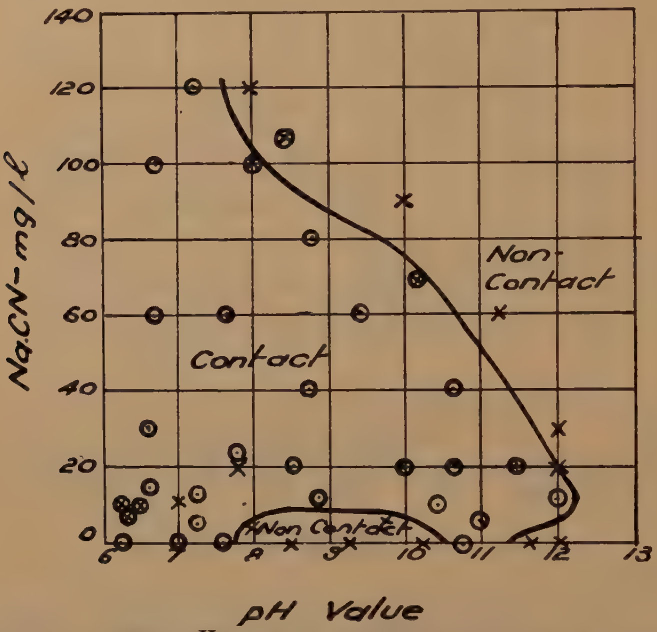

Fig 28—Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a chalcopyrite surface.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Unoxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

O = air-bubble contact.
⊗ = "catching" of an air-bubble.
X = no air-bubble contact.

Discussion of Results

It will be appreciated that in systems as complex as those studied here, it will be almost impossible to interpret many of the data with our limited physico-chemical knowledge of the constitution of the solutions. Moreover, there is no guarantee that all the curves represent equilibrium conditions. They do, however, represent what would happen in a flotation circuit of the simplest kind when a conditioning period of half an hour at 35°C was used. In the laboratory experiments, the area of mineral is very small so that collector consumption is negligible, and only the primary condition for flotation, viz., air-mineral adhesion, is considered. Many other factors are important in actual flotation and these must be considered before a successful process can be devised from the laboratory data. Despite this limitation on laboratory data, the more complex systems studied indicate how far it is possible to modify the flotation liquor, particularly if copper salts are present. Doubtless many processes which are peculiar to certain mines owe their success to the presence in the feed water of soluble salts, either originally present or else due to some soluble component of the ore. These soluble salts may be other than those usually recognized as activators.

SUMMARY

1. Oxidation of the sulphide minerals, galena, pyrite, chalcopyrite, sphalerite, produces soluble cations and sulphate in a neutral or acid pulp, and polythionates, sulphate, sulphite and thiosulphate in an alkaline pulp.

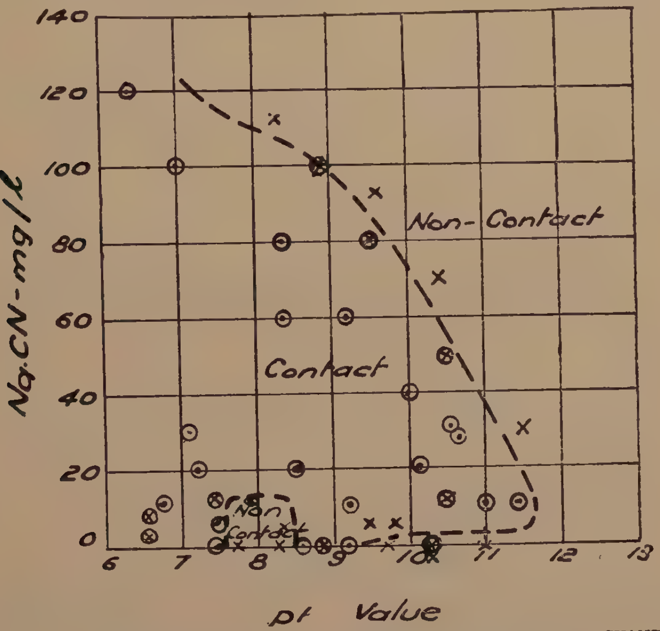

FIG 29—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT A CHALCOPYRITE SURFACE.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Oxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

O = air-bubble contact.
⊗ = "catching" of an air-bubble.
X = no air-bubble contact.

2. The amount of soluble product formed by aeration of a suspension of a sulphide mineral is governed by: (*a*) pH value of pulp; (*b*) speed of oxygen supply to mineral surface; and (*c*) time of aeration.

The following factors had little or no influence on the amount of oxidation products formed within the range studied: (*a*) solid/liquid ratio; (*b*) size of mineral; (*c*) a monomolecular film on the mineral surface; and (*d*) temperature of the pulp.

3. Ferrous and ferric salts both activate the flotation of the sulphides by ethyl xanthate. In conjunction with a copper salt and cyanide, ferrous salts improve the conditions under which it is possible to separate sphalerite from pyrite. A ferric salt in conjunction with cyanide improves the conditions for separation of chalcopyrite and galena from pyrite.

4. The anions produced by oxidation have little or no effect on flotation of fresh minerals, except that under certain conditions sulphite has a beneficial effect on the separation of chalcopyrite from galena and pyrite.

5. "Oxidizing" a sulphide mineral changes the response of that mineral to ethyl xanthate. This leads to: (*a*) greater

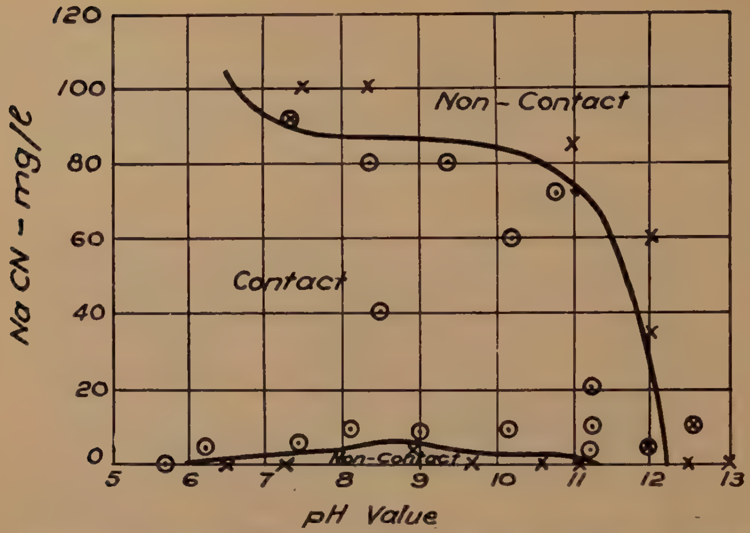

FIG 30—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT A SPHALERITE SURFACE.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Unoxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

O = air-bubble contact.
⊗ = "catching" of an air-bubble.
X = no air-bubble contact.

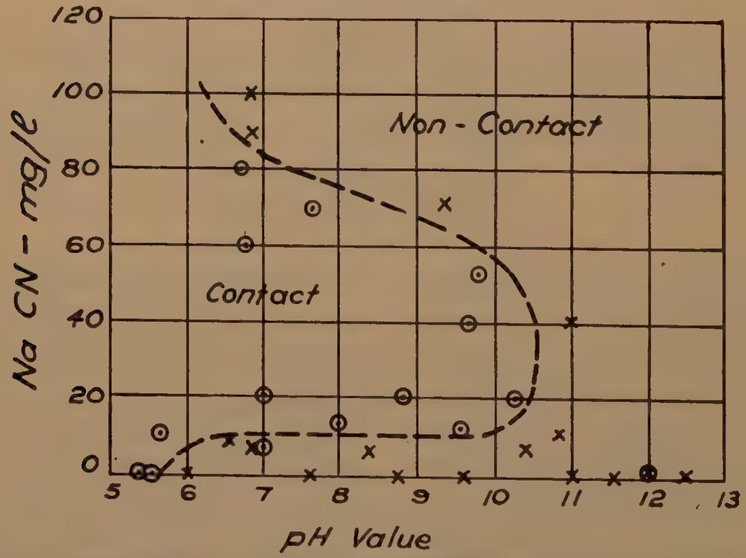

FIG 31—RELATIONSHIP BETWEEN pH VALUE AND CONCENTRATION OF SODIUM CYANIDE NECESSARY TO PREVENT CONTACT AT A SPHALERITE SURFACE.

KEtX = 5 mg/l. $CuSO_4 \cdot 5H_2O$ = 150 mg/l. $FeSO_4 \cdot 7H_2O$ = 40 mg/l. $Na_2S \cdot 9H_2O$ = 20 mg/l. Na_2CO_3 = nil. Temperature = 35°C.

Oxidized specimens pretreated for 10 min with $Na_2S \cdot 9H_2O$ solution (20 mg/l) and then washed with water before use.

O = air-bubble contact.
⊗ = "catching" of an air-bubble.
X = no air-bubble contact.

tolerance to the depressant action of cyanide and alkalis; (*b*) difficulty in separation of oxdized chalcopyrite from oxidized pyrite; and (*c*) impossiblity of separating a mixture of oxidized and unoxidized chalcopyrite from a mixture of oxidized and unoxidized pyrite.

6. With copper as an activator, oxidation improves the separation of sphalerite from pyrite provided correct reagent conditions are chosen.

7. Neither ferrous nor ferric salts improve the conditions for separation of sulphides which are partly oxidized.

8. If oxidation can be completed by using a reagent such as permanganate, separations may be more effective.

9. Pretreatment with sulphide is shown to be an effective method of ensuring more uniform behavior of a mineral which has oxidized and unoxidized surfaces.

10. Pretreatment with sulphide, followed by washing with water, leads to identical behavior of oxidized and unoxidized pyrite and galena. Oxidized chalcopyrite is partly restored to its original state and the separation of a copper-lead concentrate from a pyritic ore is made easier.

11. To avoid the necessity for removal of sulphide after pretreatment, ferrous sulphate has been added to precipitate the sulphide. Galena and chalcopyrite can then be separated from pyrite.

12. In the presence of copper, separation of partly oxidized sphalerite from partly oxidized pyrite is difficult. The separation is not improved by pretreatment with sodium sulphide.

13. Pretreatment with sulphide, followed by removal of sulphide with copper and ferrous salts does not lead to better separation of pyrite and sphalerite. The results obtained could not be satisfactorily explained.

14. Some of the solutions used in this investigation are of a degree of complexity approaching that of solutions of mill circuits. It is hoped that the results reported may be helpful to the operator in interpreting his own results and in improving his processes.

Acknowledgments

The authors wish to thank Dr. I. W. Wark for suggesting this problem and for his care in reading the manuscript.

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Flotation of Unoxidized and Oxidized Sulphide Minerals—Antimonite, Arsenopyrite, Covellite, Lollingite, Marcasite, Orpiment, Pyrrhotite and Tetrahedrite

By Enid C. Plante*

(New York Meeting, February 1948)

Introduction

To extend our knowledge of the flotation behavior of sulphide minerals, the response of the following minerals to ethyl xanthate as collector was studied by captive bubble and cylinder flotation tests: antimonite (Sb_2S_3), arsenopyrite ($FeAsS$), covellite (CuS), marcasite (FeS_2), orpiment (As_2S_3), pyrrhotite (FeS) and tetrahedrite ($4CU_2S$, Sb_2S_3). Lollingite ($FeAs_2$) was also examined to determine if its behavior could be correlated with that of pyrite or marcasite (FeS_2).

Experimental Method

Mineral for Flotation and Contact Tests

Mineral was prepared in the manner already described[1] and its behavior both freshly ground and "oxidized" was studied. Tests in which the mineral was allowed to oxidize for various periods showed that a uniform response to xanthate was reached in two days or less.

Oxidation Products

The soluble oxidation products from most of the above minerals were measured and their effect on the flotation of the mineral studied.

The mineral samples were wet ground to minus-52 mesh and deslimed, and 20 g of this material placed in 200 ml of distilled water of a measured pH value, and CO_2-free air bubbled through each suspension at a constant rate (about 1 bubble/sec) for 5 days. It has been shown[1] that the size of the mineral is unimportant in the rate of formation of oxidation products and that probably the supply of oxygen to the mineral is the rate-determining step. At the end of 5 days the pH value of the supernatant liquid was measured again, the mineral filtered off and the soluble products determined.

Methods for estimation of sulphur acids, ferrous and ferric iron and copper have already been described.[1] No satisfactory method was found for antimony. Arsenic was estimated as the arsenate.[2]

FLOTATION OF "UNOXIDIZED" MINERALS WITH ETHYL XANTHATE AS COLLECTOR

Fig 1 to 3, 4, 6, show the effect of alkali and cyanide as depressants for the minerals with 25 mg potassium ethyl xanthate per liter as collector at 35°C. Wark and Cox[3] have shown that with ethyl xanthate as collector, galena, pyrite and chalcopyrite show a greater tolerance to cyanide and hydroxyl ions at room temperature than at 35°C. Covellite (Fig 1), marcasite (Fig 2) and pyrrhotite (Fig 3) behave similarly. The independence of the curves for pyrrhotite at 35°C with respect to cyanide

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[1] References are at the end of the paper.

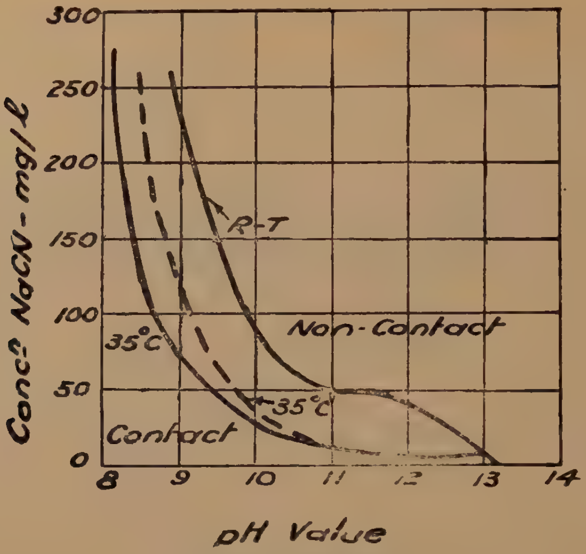

FIG 1—COVELLITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a covellite surface.

KEtX = 25 mg/l. R.T. = Room Temperature.

In this and the following curves, unless otherwise indicated:

—— = unoxidized mineral, temperature as shown.

– – – = oxidized mineral, temperature as shown.

(R.T. curve after Wark and Cox.[4])

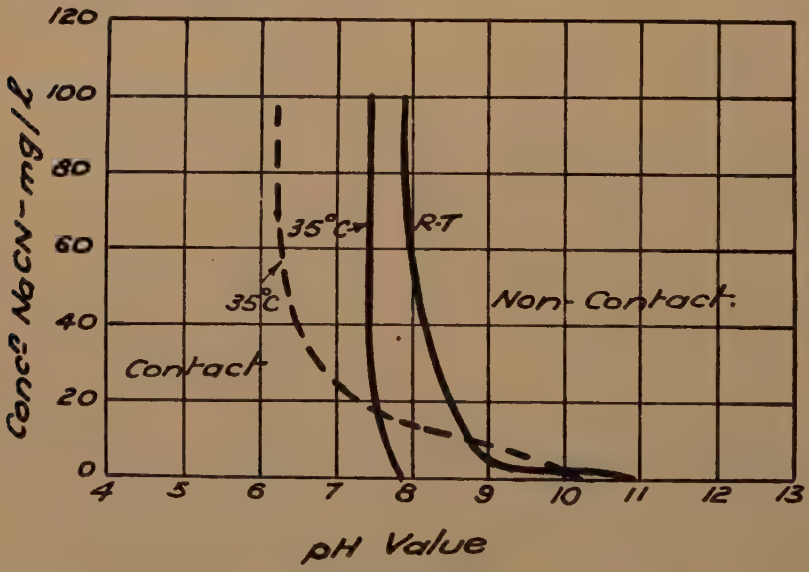

FIG 2—MARCASITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a marcasite surface.

KEtX = 25 mg/l.

(R.T. curve after Wark and Cox.[5])

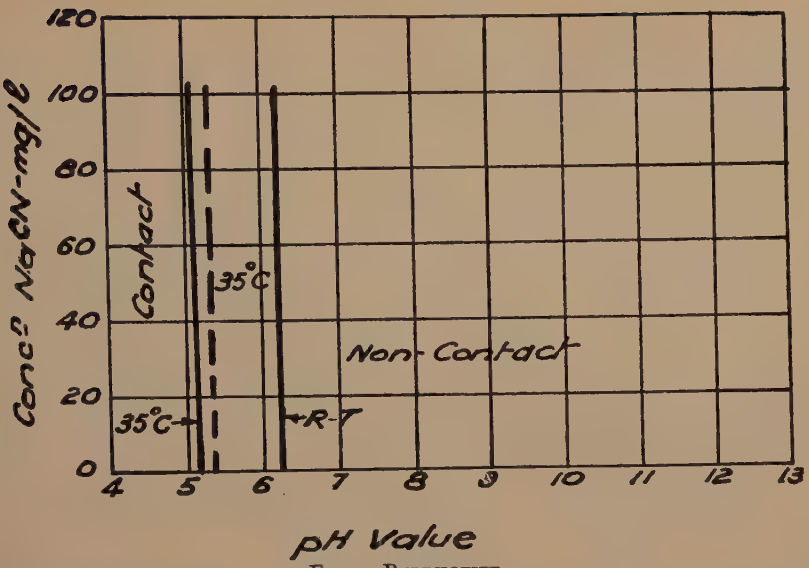

FIG 3—PYRRHOTITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a pyrrhotite surface.
KEtX = 25 mg/l.
(R.T. curve after Wark (unpublished).)

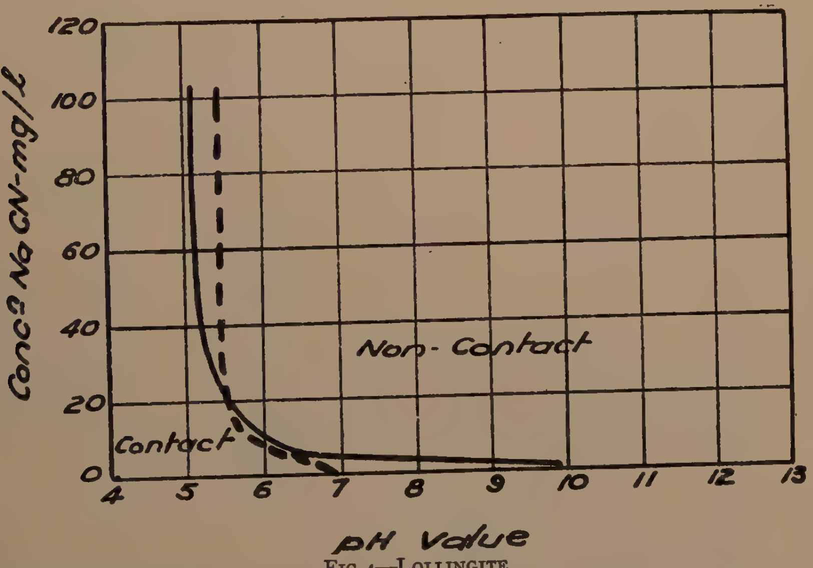

FIG 4—LOLLINGITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at lollingite surface.
KEtX = 25 mg/l.
Temperature = 35°C.

additions as high as 100 mg NaCN per liter is caused by the very small amount of cyanide ion present in solution. For an addition of 100 mg NaCN per liter the actual amount of cyanide ion in solution at pH 5 is 0.0025 mg per liter. Hence an addition of 40 g of sodium cyanide per liter would be necessary to give a concentration in solution of 1 mg CN^- per liter and this amount is probably not sufficient to cause depression. Even below pH 5, where contact is possible, the pyrrhotite floats poorly. The behavior of two minerals containing arsenic—lollingite (Fig 4) and arsenopyrite (Fig 5)—was also investigated. The results with arsenopyrite could not be closely duplicated in the absence of copper in solution. On another occasion curves fell to the left of those given in Fig 5. This lack of reproducibility is most unusual and cannot be explained. It may be related to variations in composition between the different specimens although all were mineralogically pure.

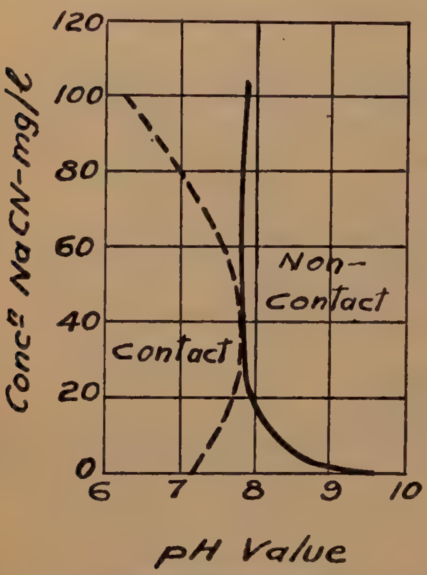

FIG 5—ARSENOPYRITE.
Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at arsenopyrite surface.
KEtX = 25 mg/l.
Temperature = 35°C.

FLOTATION OF "OXIDIZED" MINERALS

Exposure of a cleaved or polished sulphide surface to aerated water for two days enables pyrite, galena and chalcopyrite to tolerate higher concentrations of cyanide and alkali[1] for a given collector concentration. Oxidation also increases the tolerance of covellite (Fig 1), pyrrhotite (Fig 3), and tetrahedrite (Fig 6) to both cyanide and alkali.

Results for marcasite are shown in Fig 2. Lollingite (Fig 4), which will be considered with the sulphides as its flotation behavior is similar, shows decreased tolerance to alkali and slightly increased tolerance to cyanide when oxidized. Antimonite and orpiment do not float either in the unoxidized or oxidized state, with xanthate alone as collector.

Oxidation may affect flotation in two ways: (1) by alteration of the mineral surface with a consequent effect on adsorption of xanthate, and (2) by formation of oxidation products which may have depressant or activating effects. For most sulphides, particularly those containing copper, it is difficult to decide whether activation following oxidation is caused by alteration of the mineral surface or by the effect or soluble oxidation products. It has been shown[1] that anionic oxidation products have no effect on the flotation of galena, chalcopyrite, pyrite or sphalerite. The cations Cu^{++}, Fe^{++}, Pb^{++}, Ag^{+}, in general cause activation but they are probably not formed in sufficient quantity—with the exception of Fe^{++} from marcasite and Cu^{++} from covellite (Table 1) to have any effect.

EFFECT OF SULPHIDE ION

Wark and Cox have shown[6] that sulphide ion in solution depresses "unoxidized" sulphide minerals with ethyl xanthate as collector. Pretreatment of an unoxidized sulphide with sodium sulphide solution,

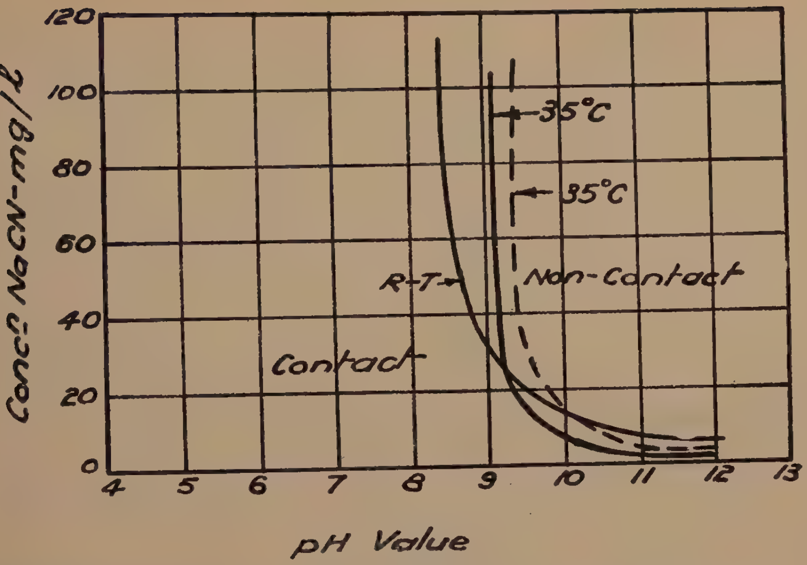

FIG 6—TETRAHEDRITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at tetrahedrite surface.

KEtX = 25 mg/l.

(R.T. curve after Wark (unpublished).)

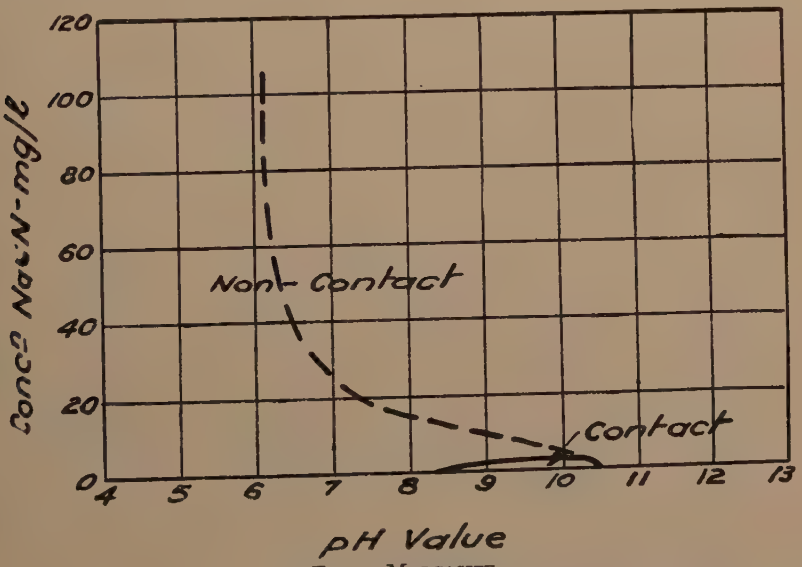

FIG 7—MARCASITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a marcasite surface.

KEtX = 25 mg/l.

Mineral pretreated 10 min in 20 mg $Na_2S \cdot 9H_2O$ per liter and then washed with water before use.

Temperature = 35°C.

——— = oxidized mineral, pretreated with Na_2S.

- - - = oxidized mineral, not pretreated (dotted curve Fig 2).

followed by washing of the mineral with water, had no effect on subsequent flotation.[1] On the other hand, pretreatment of oxidized sulphides with sulphide solution restored the behavior of galena and pyrite to that of the unoxidized mineral, whereas the behavior of oxidized chalcopyrite pretreated with sulphide was intermediate between that of oxidized and unoxidized mineral. Pretreatment of oxidized marcasite with sodium sulphide however had a strongly depressant effect on its flotation (Fig 7), there being a small region of flotation only between pH of 8 to 10 and at cyanide concentrations below 5 mg per liter. This depression is not caused by residual sulphide ion in solution since similar treatment of unoxidized mineral produced no depression, indicating that all sulphide has been washed away.

Copper Sulphate as an Activator

Carbonate Present in Solution

Fig 8 and 9, and 12 to 15 show the response to 5 mg potassium ethyl xanthate per liter of antimonite, arsenopyrite, marcasite orpiment, tetrahedrite and lollingite in the presence of 150 mg per liter of both copper sulphate and sodium carbonate. All tests were at 35°C. Oxidation of galena, pyrite, chalcopyrite and sphalerite decreases the tolerance of the mineral to alkali and cyanide below the inflection of the curve at 70 to 90 mg per liter.[1] The initial reaction in solution between cyanide and copper sulphate is precipitation of white cuprous cyanide.[7]

$$CN^- + Cu^{++} \rightleftharpoons CuCN \downarrow + \tfrac{1}{2}C_2N_2$$

Additional NaCN dissolves the precipitate to form a soluble cuprocyanide:

$$CN^- + CuCN \rightleftharpoons [Cu(CN)_2]^-$$

Thus when the molecular ratio of NaCN to copper sulphate reaches 3:1, i.e., theoretically at a concentration of 88 mg per liter of NaCN for a 150 mg per liter solution of $CuSO_4 \cdot 5H_2O$, all the copper in solution is converted into the complex ion $Cu(CN)_2^-$.

Until the addition of NaCN reaches a value which is stoichiometrically equivalent to the soluble copper sulphate for formation of the complex, the cyanide-ion concentration in solution is very low, but as soon as this value is exceeded, further addition of sodium cyanide raises the concentration of cyanide ion in solution rapidly with consequent depression of the mineral. The mineral may be depressed at concentrations other than 88 mg per liter of NaCN because of the complexity of the system—e.g., precipitation of some of the copper by hydroxyl or carbonate, or the possible formation of other copper cyanide complexes which may depress. Wark[8] suggests that the cuprocyanide ion may depress pyrite under some conditions.

Antimonite (Fig 8) when oxidized behaves the same as the unoxidized mineral at cyanide concentrations above 75 mg per liter but below this there is greater depression of the oxidized mineral. There is a region of poor flotation at about 20 mg per liter of NaCN and pH of 6 to 7. Wark and Sutherland[9] have shown that with low xanthate concentration (5 mg per liter) and sodium carbonate present in solution with copper sulphate, there is often an island of nonflotation centered about 30 mg per liter of NaCN and pH of 6.5. Unoxidized *arsenopyrite* exhibits a large island of this type (Fig 9). The oxidized mineral shows a lower tolerance to cyanide and alkali throughout the whole range which is unusual, and the boundary of the island of nonflotation has merged into that of the main region of nonflotation to give a continuous curve, with the shape of the island still maintained. Depression of the oxidized mineral might be caused by arsenic oxidation products. Table 1 shows that arsenic is produced in appreciable quantity when arsenopyrite is oxidized. Fig 10 and 11 show that arsenite and arsenate have a depressant effect on the flotation of arsenopyrite in the absence of

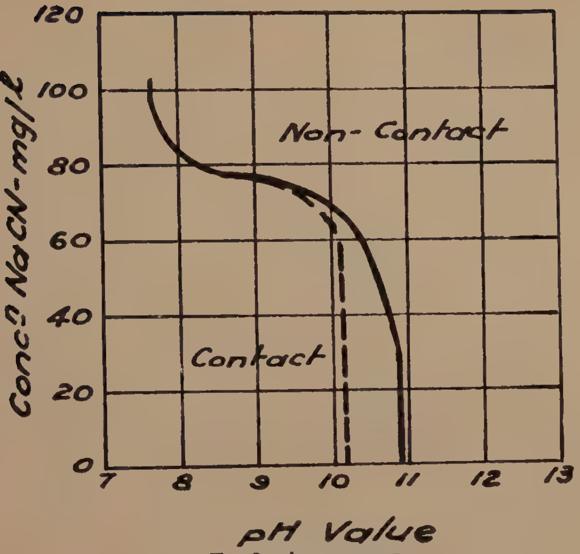

FIG 8—ANTIMONITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at an antimonite surface.

KEtX = 5 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.

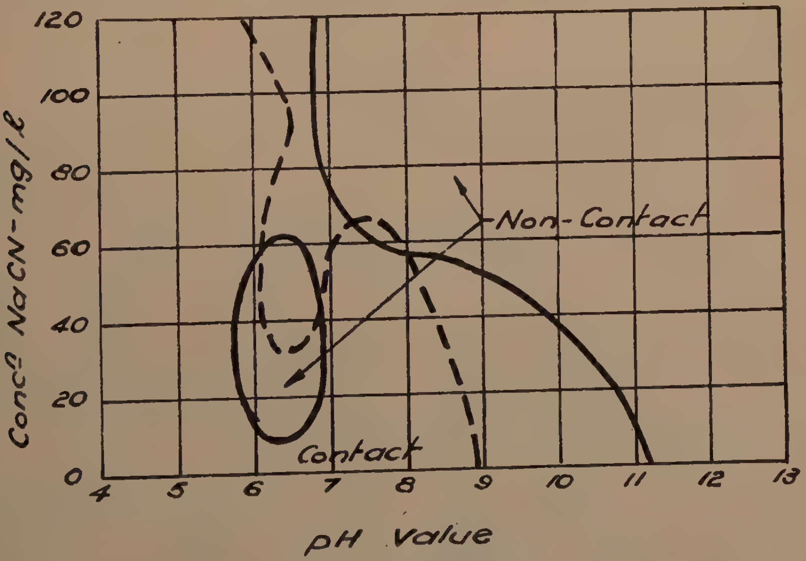

FIG 9—ARSENOPYRITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at an arsenopyrite surface.

KEtX = 5 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.
(Unoxidized curve after Wark and Sutherland.[9])

copper. However these results should be considered only to show the general trend of the behavior of arsenopyrite as some doubt exists about the reproducibility of results.

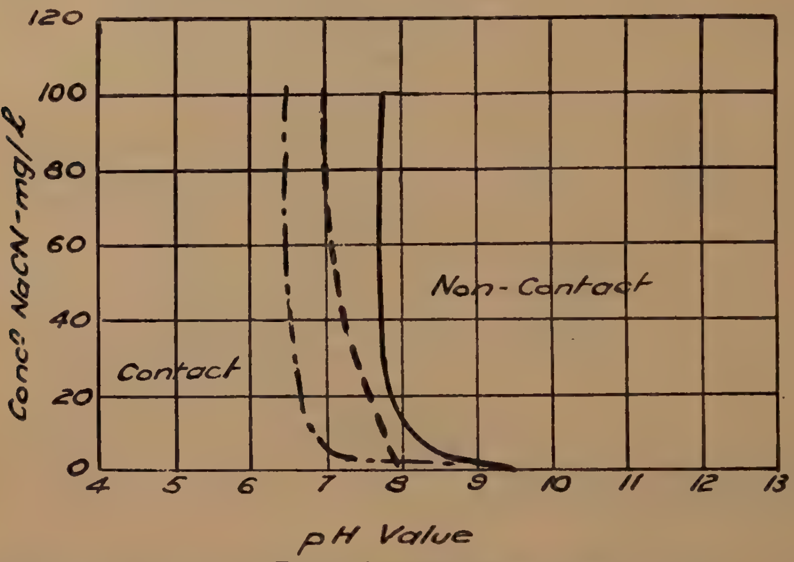

FIG 10—ARSENOPYRITE.

Relationship between pH value and concentration of sodium cyanide necessarv to prevent contact at an arsenopyrite surface.

KEtX = 25 mg/l.
Temperature = 35°C.
—— = unoxidized, no other addition.
– – – = unoxidized, 50 mg potassium arsenite per liter.
–.–. = unoxidized, 50 mg As_2O_5 per liter.

Marcasite (Fig 12) when unoxidized shows a greater sensitivity to cyanide than do the other minerals but the curve is of the normal type. Wark and Sutherland[9] have shown (their Fig 2) successive stages in departure from the standard type of curve as depressant conditions are intensified with xanthate type of collector and in the presence of a copper salt. The curve for marcasite is like Type *D* of this series and is like the curve for pyrite under the same conditions except that instead of 150 mg per liter there is 1500 mg per liter of sodium carbonate in their solution (their Fig 6). Oxidized marcasite does not float under these conditions and thus differs from the other sulphides, notably from pyrite which has the same composition.

Orpiment (Fig 13). The curves for oxidized and unoxidized surfaces cross at a cyanide concentration of about 30 mg per liter.

Tetrahedrite. (Fig 14) behaves normally, exhibiting a high tolerance to cyanide when unoxidized, the critical pH value being above 9.5 at 120 mg per liter of NaCN. Even in the absence of copper it is high (> 9) (Fig 6). When oxidized, the mineral shows a lower tolerance to cyanide and alkali ion, and there is a small island of noncontact about pH of 6.5 and 30 mg of NaCN per liter.

Lollingite (Fig 15) when unoxidized behaves similarly to arsenopyrite and there is an island of nonflotation about pH of 6.5 and 40 mg per liter of NaCN. Oxidation, as with arsenopyrite, depresses the lollingite but to a more marked degree and the maximum cyanide concentration at which flotation occurs is 35 mg per liter at pH of 7.5 to

8. On either side of this pH value the region of flotation increases with decreasing cyanide concentration. There is a region of nonflotation for cyanide at 0 to 4 mg per liter and at pH of 8.4 to 10. Whether these effects are partly or wholly caused by oxidation products is unknown. It is not known how oxidation of the lollingite ($FeAS_2$) proceeds—analyses of its oxidation products have shown that sulphur acids are formed (Table 1) so that there are sulphur-containing impurities in the mineral used. Soluble anionic sulphur-containing oxidation products formed from the sulphides are unlikely to affect flotation[1] but it could be affected by anionic arsenic products, e.g., arsenate, or by soluble cationic oxidation products. Table 1 shows that very little arsenate or iron is formed on oxidation of the lollingite, and the depression is probably related to the presence of copper in solution as it is not nearly so marked in the absence of copper (Fig 4).

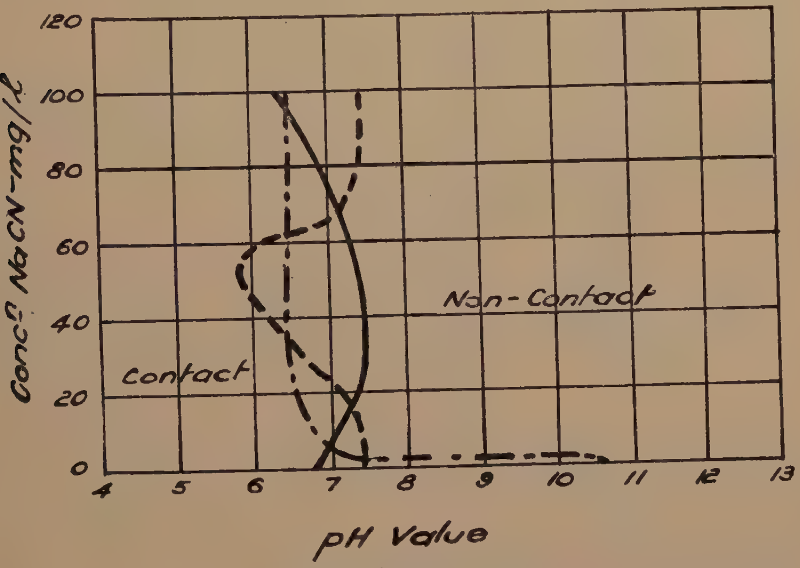

FIG 11—ARSENOPYRITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at an arsenopyrite surface.

KEtX = 25 mg/l.

Temperature = 35°C.

——— = oxidized, no other addition.

- - - = oxidized, 50 mg potassium arsenite per liter.

-.-. = oxidized, 50 mg As_2O_5 per liter.

Absence of Carbonate

The effect of omitting sodium carbonate from the above system is seen for marcasite in Fig 16. The tolerance of the unoxidized mineral to cyanide is much increased and the curve is similar in shape to that for oxidized arsenopyrite in the presence of Na_2CO_3, $CuSO_4$ and low xanthate (Fig 9), in which the island of nonflotation has merged into the general region of nonflotation, to give a continuous curve. It is unusual to get islands of nonflotation in the absence of carbonate at 35°C. They are more usual with high Na_2CO_3 and low xanthate when copper is present. Wark and Sutherland[9] have shown that carbonate ion influences the behavior of sulphides in the presence of copper because it forms basic carbonate precipitates with the copper. This may result in less cyanide being tied up as complex copper cyanide ions, and so more free cyanide is available in solution at lower cyanide additions, which accounts for the decreased tolerance to cyanide in the presence of carbonate. The oxidized mineral also floats at pH values below 6.

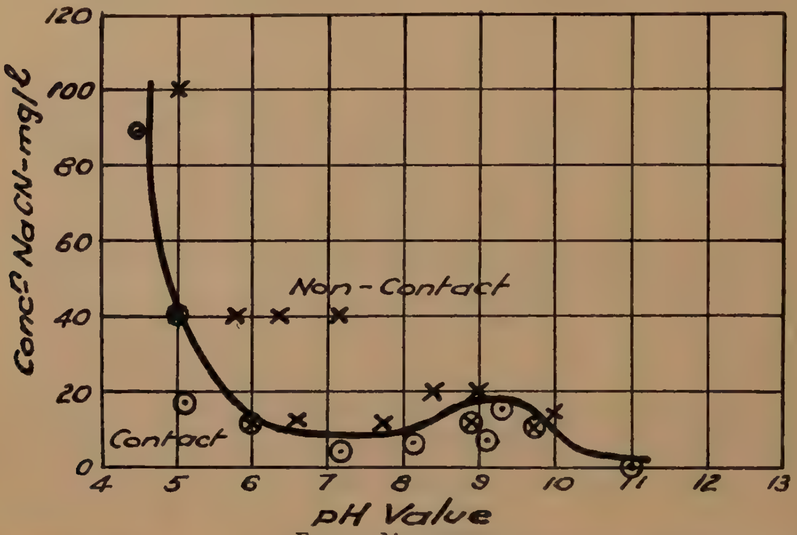

FIG 12—MARCASITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a marcasite surface.

KEtX = 5 mg/l.
$CuSO_4$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.
——— = unoxidized mineral.
⊙ = air-bubble contact.
⊗ = "catching" of air-bubble.
X = noncontact of air-bubble.

XANTHATE CONCENTRATION

With 150 mg of both copper sulphate and sodium carbonate per liter in solution, the effect of xanthate concentration on the flotation of arsenopyrite is seen in Fig 17. With 25 mg of potassium ethyl xanthate per liter the flotation region is depressed well below that for 5 mg per liter. One can see how the curve for 25 mg of xanthate per liter is developed from that for 5 mg per liter, the region of nonflotation of the island and the general region merging into one another and the area below 50 mg of NaCN per liter becoming one of nonflotation. This depression by higher xanthate is very unusual; usually the higher xanthate increases the tolerance to cyanide and alkali ions.[1] Although there is some doubt about the validity of results for arsenopyrite, these results have been checked by contact and cylinder tests and the effect seems to be a real one.

The flotation curves for unoxidized and oxidized covellite under the same conditions are shown in Fig 18. Oxidation depresses the mineral and decreases its tolerance to cyanide.

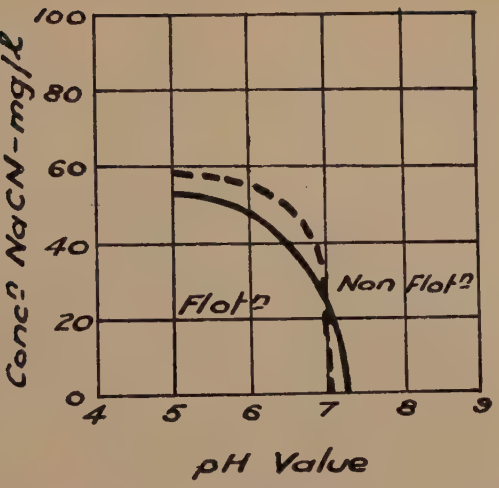

FIG 13—ORPIMENT.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at an orpiment surface.

KEtX = 5 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.

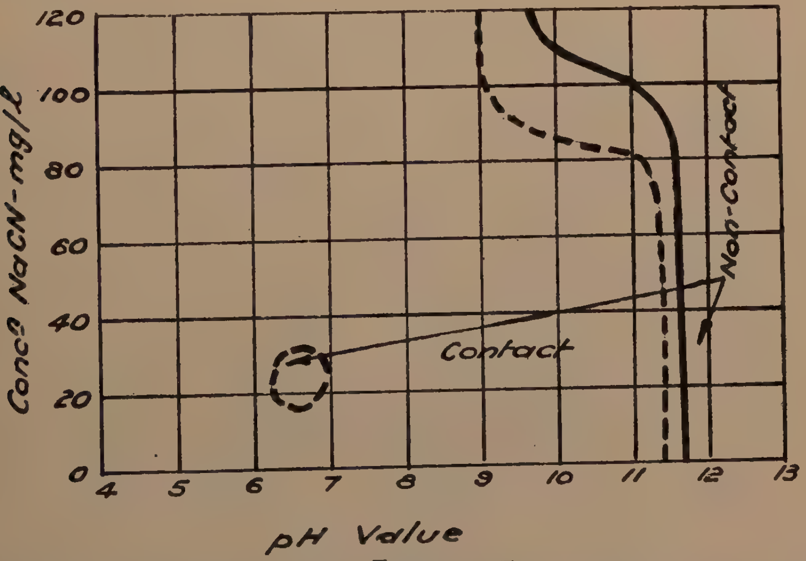

FIG 14—TETRAHEDRITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a tetrahedrite surface.

KEtX = 5 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.

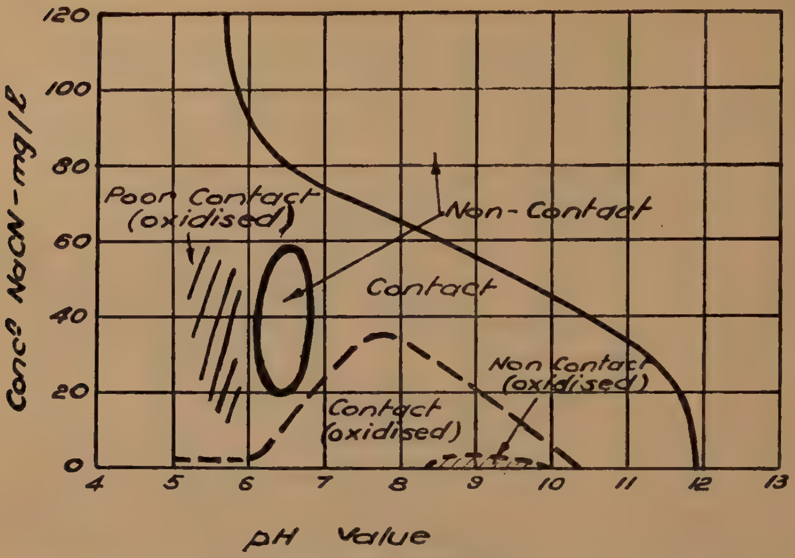

FIG 15—LOLLINGITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a lollingite surface.

KEtX = 5 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Temperature = 35°C.

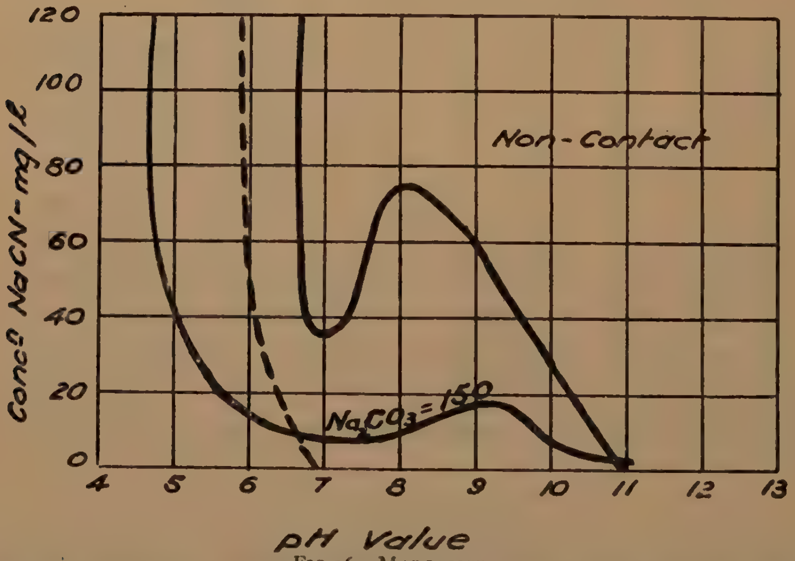

FIG 16—MARCASITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a marcasite surface.

KEtX = 5 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = Nil.
Temperature = 35°C.

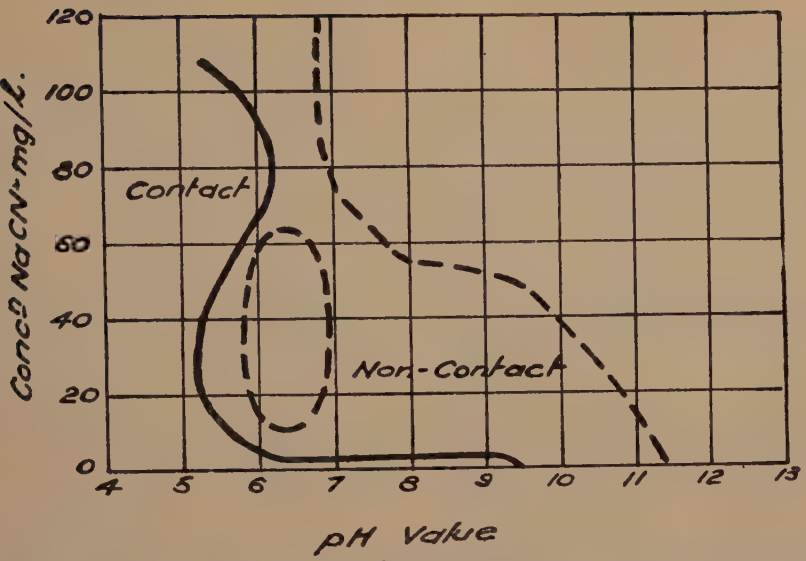

FIG 17—ARSENOPYRITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at an arsenopyrite surface.

KEtX = 25 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.
—— = unoxidized mineral. KEtX = 25 mg/l.
– – – = unoxidized mineral. KEtX = 5 mg/l. (from Fig 9).

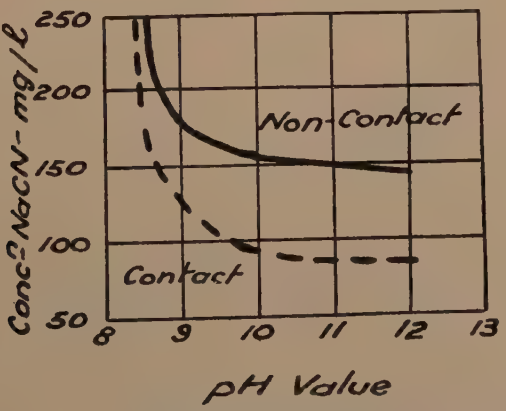

FIG 18—COVELLITE.

Relationship between pH value and concentration of sodium cyanide necessary to prevent contact at a covellite surface.

KEtX = 25 mg/l.
$CuSO_4 \cdot 5H_2O$ = 150 mg/l.
Na_2CO_3 = 150 mg/l.
Temperature = 35°C.

OXIDATION PRODUCTS

Table 1 shows the soluble oxidation products found after oxidation of the minerals as described on page 189. Two suspensions of each mineral were oxidized, one with the pulp starting close to the neutral pH value 7 and the other at the alkaline pH value of 10.7.

The pH values of the solutions after oxidation varied considerably, e.g., antimonite and tetrahedrite maintained values above 7 for both solutions, whereas the marcasite solutions dropped rapidly to very low values. However, the much smaller amount of ferrous iron resulting from the "alkaline" pulp indicates that the pH of this pulp was probably higher than that of the "acid" one for most of the oxidation period. In all solutions the amount of sulphur acids formed was greater when the initial pH value was high. This was also found when galena and others were oxidized.[1]

TABLE 1—*Soluble Oxidation Products*

Mineral	pH Value of Liquid		Oxidation Product. mg per liter				
	At Start of Oxidation	After 5 Days	Fe^{+++}	Fe^{++}	Cu^{++}	Arsenic (as Arsenate)	Total SO_4^{-}
Antimonite	6.8	7.5					45
Antimonite	10.7	9.2					50
Arsenopyrite	6.8	5.7				7–10	50
Arsenopyrite	10.7	6.9				5–8	80
Covellite	6.8	5.7			11		20
Covellite	10.7	6.9			2		60
Lollingite	6.8	4.1		2		1	20
Lollingite	10.7	7.1		1			50
Marcasite	6.8	2.5	3	49			180
Marcasite	10.7	3.5		7			190
Orpiment	6.8	3.9					35
Orpiment	10.7	5.5					160
Pyrrhotite	6.8	5.1		6			80
Pyrrhotite	10.7	10.7		1			110
Tetrahedrite	6.8	7.1					20
Tetrahedrite	10.7	8.4					135

None of the anionic sulphur oxidation products affect flotation of minerals studied previously[1] so it is unlikely that they will affect flotation of the minerals studied here.

Of the cationic products, the ferrous iron from marcasite at pH of 6.8 to 2.5 (49 mg per liter) may have a small activating effect, the 11 mg per liter of Cu^{++} from covellite will activate and the arsenic from arsenopyrite may have a depressant effect. The smaller amounts of ferrous iron, ferric iron and copper—all less than 10 mg per liter—are probably insufficient to have any noticeable effect on flotation.

SUMMARY

1. The flotation of antimonite, arsenopyrite, covellite, lollingite, marcasite, orpiment, pyrrhotite and tetrahedrite has been studied. The soluble oxidation products formed from these minerals were also investigated.

2. In general, oxidation increases the tolerance of the minerals to cyanide and alkali ion with ethyl xanthate as collector. The activation caused by oxidation is probably due to alteration of the mineral surface and its effect on xanthate adsorption rather than to the effect on the mineral of the oxidation products formed.

3. Pretreatment of oxidized marcasite by sodium sulphide solution strongly depresses the mineral.

4. These minerals when unoxidized are activated by the presence of copper and carbonate in solution. Oxidation depresses

antimonite, arsenopyrite, marcasite, tetrahedrite and lollingite in the presence of copper and carbonate.

5. Omission of carbonate from the system increases the tolerance of marcasite to cyanide and alkali ions.

6. Raising the xanthate concentration in the presence of copper and carbonate depresses unoxidized arsenopyrite.

7. Oxidation of the minerals produces mainly thio-acids which have been estimated as sulphate. It is unlikely that these thio-acids will affect the flotation of the minerals.[1] Soluble arsenic compounds are formed when arsenopyrite is oxidized and may cause depression. The only cations produced in appreciable amount are ferrous iron from marcasite and copper from covellite. The copper will activate the covellite and the ferrous iron will probably activate also.

ACKNOWLEDGMENTS

The author gratefully acknowledges the assistance of Misses M. Collins and B. Wilson in the experimental work.

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The Determination of Oxide Lead in Ores and Concentrator Products

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(New York Meeting, March 1947)

The differentiation of oxidized forms of lead from lead sulphide in complex products by chemical analysis is of considerable importance to certain mining and metallurgical companies. A method for the separation of "oxide" lead, (non-sulphide), from sulphide lead, based on the solubility of the former and virtual insolubility of the latter in concentrated ammonium acetate solution, has been employed for many years in the industry.[1,2] Unfortunately this solvent, while satisfactory for lead sulphate and carbonate, is without action on lead phosphates or vanadates, with the result that at Rhodesia Broken Hill Development Co. where these latter minerals are present, they have counted as sulphides when using this procedure. For the control of operations in the new concentrator it became necessary to have a closer check on sulphide lead than could be obtained by the only current method of differentiating this constituent from oxidized lead.

For some time this problem has engaged our attention, and after a large number of solvents had been tried, the following procedure was evolved. Like all methods depending on selective solution in a complex material, it will not give 100 pct separation into oxide and sulphide portions. Because of the ease of oxidation of galena this is almost an impossibility. Our procedure will, however, give a recovery of 99–100 pct of the oxidized lead, contaminated with very small quantities of lead sulphide, and a corresponding recovery of 97–98 pct of the sulphide lead. The procedure is simple and reproducible, and gives results quite satisfactory for routine control purposes.

Experimental

For this work standard samples of the following pure lead minerals were obtained from Broken Hill, Northern Rhodesia. These were hand-picked specimens, and were carefully ground to −200 mesh. Since it was realized that a fair-sized sample would be necessary for this work to thoroughly test the large number of procedures we had in mind, a compromise had to be effected between a few grams of very pure minerals and a large sample of lower grade material. For this work the presence of gangue was not deleterious, so long as the oxide lead minerals did not contain lead sulphide, and vice versa.

Table 1 shows the lead minerals used, and their total lead content, both theoretical and actual. Apart from the pyromorphite the minerals are seen by analysis to be quite pure.

Many solutions were tried in an effort to dissolve the oxidized lead minerals and

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[1] References are at the end of the paper.

leave the sulphide intact, or to selectively attack sulphides. Among these might be mentioned dilute HCl, NH_4Cl, dilute sulphuric acid saturated with SO_2 in presence of ammonium acetate, dilute NaOH, H_3PO_4, NH_4NO_3, NaCl + $FeCl_3$, dilute HF, NH_4OH, dilute $FeSO_4$, KOH + $(NH_4)C_2H_3O_2$, $NH_4Cl + (NH_4)C_2H_3O_2$, $HClO_4$. Several variations in reagent concentration, time of contact, and temperature were employed for the more promising solvents, but it was invariably found that a solvent powerful enough to completely dissolve the oxidized lead minerals attacked the sulphides to an extent sufficient to condemn the procedure.

TABLE 1—*Composition of Lead Minerals Employed*

Mineral	Formula	Percentage Lead	
		Theoretical	Actual
Galena	PbS	86.6	81.60
Anglesite	$PbSO_4$	68.3	67.73
Cerussite	$PbCO_3$	77.5	73.23
Pyromorphite	$3Pb_3P_2O_8 \cdot PbCl_2$	76.4	57.12
Descloizite	$2PbO \cdot 2ZnO \cdot V_2O_5 \cdot H_2C$	51.2	48.18
Vanadinite	$Pb_3V_2O_8 \cdot PbCl_2$	73.4	69.35

Finally, the solubility of anglesite and cerussite and insolubility of other forms of lead in ammonium acetate solution, and the relative insolubility of lead sulphide coupled with the solubility of lead phosphate and vanadates which we found in dilute perchloric acid solution, suggested a promising procedure. The latter consists of first leaching with ammonium acetate when lead sulphate and lead carbonate, together with any lead oxide if present, will be dissolved, leaving the other forms of lead in the residue. Treatment of the latter with 10 pct perchloric acid in the cold under the conditions outlined below will dissolve pyromorphite, descloizite, and vanadinite, while leaving lead sulphide relatively unattacked. Thus a good differentiation between sulphide and oxidized forms of lead is secured.

Dilute perchloric acid was found to be a good solvent for oxidized lead minerals, the high solubility of lead perchlorates aiding the reaction. Cold perchloric acid, of course, is such a weak oxidizing agent that it does not even liberate iodine from iodides dissolved in it, and for this reason has little action on sulphides.

PROCEDURE

The procedure finally adopted was as follows:

Weigh out 1 g of −200 mesh sample into a small stoppered Erlenmeyer flask and add 100 ml of 60 pct ammonium acetate solution. Allow to stand at room temperature for 1½ hr, shaking the flask every 10–15 min. Filter through a 15 cm No. 42 Whatman paper and wash by decantation thoroughly with cold water, retaining as much of the residue in the beaker as possible. A small quantity of material will unavoidably pass onto the filter paper, but this should be kept to a minimum. The filtrate contains lead present as sulphate, carbonate, and oxide.

Replace the original flask under the funnel and pour through the paper 100 ml of cold 10 pct perchloric acid.* Reserve any lead sulphide residue on the paper for addition to final sulphide residue. Stopper the flask and allow to stand at room temperature for 2 hr, shaking the flask for a few seconds every 10–15 min. Filter through a 15 cm No. 42 Whatman filter paper having an 11 cm No. 42 paper as a cone outside the former, and wash thoroughly with cold 1 pct perchloric acid. The remaining oxide lead minerals like pyromorphite, descloizite, and vanadinite are in the filtrate, while lead sulphide remains on the filter paper.

Owing to the tendency for finely-divided lead sulphide to oxidize on exposure to air and moisture, the filtrations should be carried through as rapidly as possible, and prolonged washing should be avoided. The presence of laboratory fumes increases this tendency of lead sulphide to oxidize.

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*10 pct by volume concentrated perchloric acid, 72 pct.

If the lead minerals are very finely disseminated in a siliceous gangue it may be necessary to add a few drops of HF to the sample during both leaches. Small quantities of HF are without effect on lead sulphide, as high as 5 pct HF dissolving insignificant amounts under the conditions of the above procedure.

The oxide lead in the filtrates from the ammonium acetate and perchloric acid treatments may be combined and lead determined by any suitable procedure. They may also, if desired, be determined separately, thus differentiating between lead present as sulphate or carbonate and lead found as phosphate or the vanadates. The final residue of lead sulphide may be easily dissolved in nitric acid and bromine, and lead determined as usual.

Owing to the time required to decompose the finely-divided mineral particles carried into the filter paper following the perchloric acid treatment, it will be found preferable to determine sulphide lead from the relation: total lead − oxide lead = sulphide lead.

Results

The pure minerals were first investigated singly, and Table 2 illustrates the results obtained:

Table 2—*Extraction of Lead by Ammonium Acetate and Perchloric Acid, Respectively*

Mineral	Lead Present, Grams	Lead Extracted by Ammonium Acetate		Lead Extracted by Perchloric Acid	
		Grams	Per Cent	Grams	Per Cent
Galena	0.8160	0.0037	0.45	0.0175	2.14
Anglesite	0.6773	0.6772	100		
Cerussite	0.7323	0.7321	100		
Pyromorphite	0.5712	0.0047	0.82	0.5711	100
Descloizite	0.4818	0.0026	0.54	0.4794	99.5
Vanadinite	0.6935	0.0035	0.50	0.6876	99.1

The small quantities of lead extracted from galena, pyromorphite, descloizite, and vanadinite by ammonium acetate are undoubtedly in large part the result of traces of lead sulphate and carbonate present on these minerals. This is indicated by the fact that the quantity of lead extracted remained substantially the same with variations in time of contact, temperature, and concentration of ammonium acetate. Even if some of this were caused by inherent solubility of these minerals in ammonium acetate, however, the percentage of the total lead extracted, ranging from 0.45 to 0.82 pct, is low enough to be ignored in works practice.

Perchloric acid treatment of the mixed sulphide-oxide residue from the ammonium acetate leach results in 99–100 pct extraction of the oxide portion, with the simultaneous solution of about 2 pct of the lead present as sulphide. This degree of accuracy is almost within that allowed for sampling and analytical variations and should be quite satisfactory for control purposes.

After thoroughly testing out the procedure on individual mineral types a composite sample was prepared by weighing out equal quantities of 0.25 g of all the Broken Hill lead minerals, mixing, and carrying out the procedure outlined above. The results are shown in Table 3.

Table 3—*Extraction of Lead Minerals in a Composite Oxide-sulphide Sample*

Sulphide Lead			Oxidized Lead		
Present, Gram	Found, Gram	Extraction, Per Cent	Present, Gram	Found, Gram	Extraction, Per Cent
0.2040	0.1987	97.4	0.7891	0.7920	100.4

Of course, the solubility of lead sulphide in the solvents and consequent contamination of the oxide filtrates with a little lead derived from the sulphide explains the slightly higher quantity of oxide lead found than was actually present. Subtracting the theoretical quantity of lead, dissolved out of the sulphide, from the oxide lead found, the extraction of the latter would be 99.7 instead of 100.4 pct.

This method when applied to complex concentrator products gives an excellent agreement with the mineral differentiation found by microscopic studies.

SUMMARY

A procedure has been developed for differentiating by chemical analysis oxidized lead from sulphide lead in ores and concentrator products. This method overcomes the disadvantage of existing procedures wherein lead phosphates and vanadates are counted as sulphides.

The method consists in leaching first with ammonium acetate solution to dissolve sulphate, carbonate, and oxide, and extracting the residue with dilute perchloric acid. Lead phosphate and vanadates are soluble in the latter reagent, while lead sulphide is only attacked to a very slight degree under the recommended conditions.

The procedure is simple, reproducible, and quite satisfactory for control purposes. It gives a recovery of 99–100 pct of the oxide lead and 97–98 pct of the sulphide lead in ores and concentrator products.

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Attachment of Mineral Particles to Air Bubbles in Flotation

By H. Rush Spedden,* Member AIME and William S. Hannan, Jr.†

(New York Meeting, February 1948)

FLOTATION may be defined as a process whereby mineral particles are concentrated by selective adhesion to air-liquid interfaces. The process involves attachment of desired mineral particles to air bubbles, and removal of the mineralized air bubbles from the surface of the pulp as a mineral concentrate.

PROBLEM

Although great strides have been made in understanding the principles of flotation, one part of the process which has not been adequately explained is the mechanism by which selected mineral particles become attached to air bubbles. A glance at the froth of an operating flotation cell demonstrates adequately that use of proper reagents and conditions will result in such attachments; but how such attachment takes place is not definitely known.

A better knowledge of this factor might lead to improved flotation practice either through altered machine design or the application of scientific innovations.

Direct observation of the behavior of mineral particles and bubbles in an operating cell does not appear promising because of the complexity of the system. For this reason, conclusions concerning the problem must be deduced from theoretical considerations or from observation of simplified systems designed to reproduce conditions thought to exist in an operating cell.

PRESENT THEORIES

Two mechanisms have been recognized as fundamental in particle-bubble attachment: gas precipitation and direct encounter. It is a simple matter to demonstrate in the laboratory that attachment can be brought about by either mechanism. As to the relative importance of the two in a particular flotation system, however, there is a divergence of views. Taggart[1] states,

> Under exceptional circumstances sufficient force and time to effect attachment may conjoin in an ore pulp undergoing agitation, and a preformed bubble adhere to a collector-coated mineral particle, but the occurrence is rare. . . . Pulp-body froth flotation effects selection in the body of a conditioned pulp by causing gas bubbles to precipitate from solution in the liquid phase preferentially at the surfaces of collector-coated particles, and to cling to these particles.

Gaudin[2] states that, within the pulp body of an agitation machine, attachment of mineral particles to air bubbles occurs as a consequence of contact between the two. This theory has been supported by theoretical reasoning and much circumstantial evidence. Perhaps the most interesting bit of evidence in favor of this theory is the difficulty in the flotation of fines. If attach-

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[1] References are at the end of the paper.

ment of particles to bubbles were a result of precipitation, it would seem that fines would be at least equally as floatable as normal flotation sizes. On the other hand, if mineralization occurs by contact, at least a part of the difficulty in the flotation of fines is readily explained by a consideration of laminar-flow effect.

These two theories are here simply stated. The references cited should be consulted for more complete details.

SCOPE OF WORK

The present study was initiated in an effort to learn more about mineral-bubble contact and attachment. At first it was felt that much valuable information could be obtained by applying high-speed motion photography to a simple captive-bubble system such as used by Taggart, Taylor and Ince.[3] The plan was to observe the formation of an air-mineral interface. Early in the work it was discovered that a single captive bubble did not provide sufficient flexibility and therefore a system was developed so that observations could be made of a series of bubbles rising freely through a stream of falling mineral particles. Even though particle-bubble contacts were photographed at 3000 frames per second, the actual contact took place so rapidly that a free particle in one frame was found to be firmly attached to a bubble in the succeeding frame. Thus, although the precise details of formation of an air-mineral interface were not observed even at 3000 frames per second, some interesting and unexpected facts were discovered.

EXPERIMENTAL TECHNIQUE

In its final form the experimental apparatus consisted of a glass flotation cell mounted on the stage of a microscope with horizontal axis. Magnification of the microscope was 16 ×. Air bubbles, approximately 2 mm in diameter, were introduced at a point directly below the field of view by blowing air through a capillary tube. Xanthate-conditioned galena particles were added as a pulp to a point below the liquid surface in the cell so that a falling stream of particles would meet a rising column of air bubbles within the field of view of the microscope. Events were recorded by an Eastman high-speed camera operating at 3000 frames per second. Normal projection speed slows down these events by a factor of 187.5.

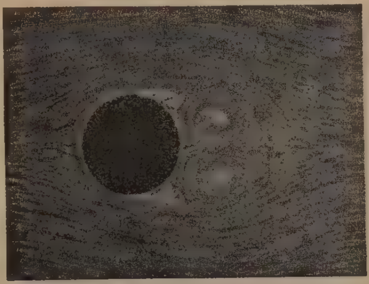

FIG 1—STATIONARY VORTEX IN WAKE OF A CYLINDER (TURBULENT FLOW).

The camera was at rest with respect to the cylinder. (After Prandtl-Tietjens—published by permission from "Principles of Mineral Dressing" by A. M. Gaudin, copyrighted, 1939, by McGraw-Hill Book Co., Inc.)

All experiments were performed with galena particles which had been ground in a porcelain mortar while immersed in a 25 mg per liter solution of potassium ethyl xanthate and then sized to minus-48, plus-200 mesh. The solution used in the cell was of the same composition.

BEHAVIOR OF MINERAL PARTICLES AND BUBBLES AS OBSERVED ON MOTION MICROGRAPHS

High-speed micrographs taken by this technique have enabled us to study the behavior of mineral particles in the vicinity of an air bubble. Many facts have been proven which were formerly surmised. In the first instance, air bubbles of the order of magnitude of one to two millimeters rise through water under conditions of turbulent flow. These conditions are well shown in Fig 1, a photograph, borrowed from another source,[8] which portrays turbulent flow in the wake of a cylinder. Similar conditions in the region of rising air bubbles have been observed in the present study.

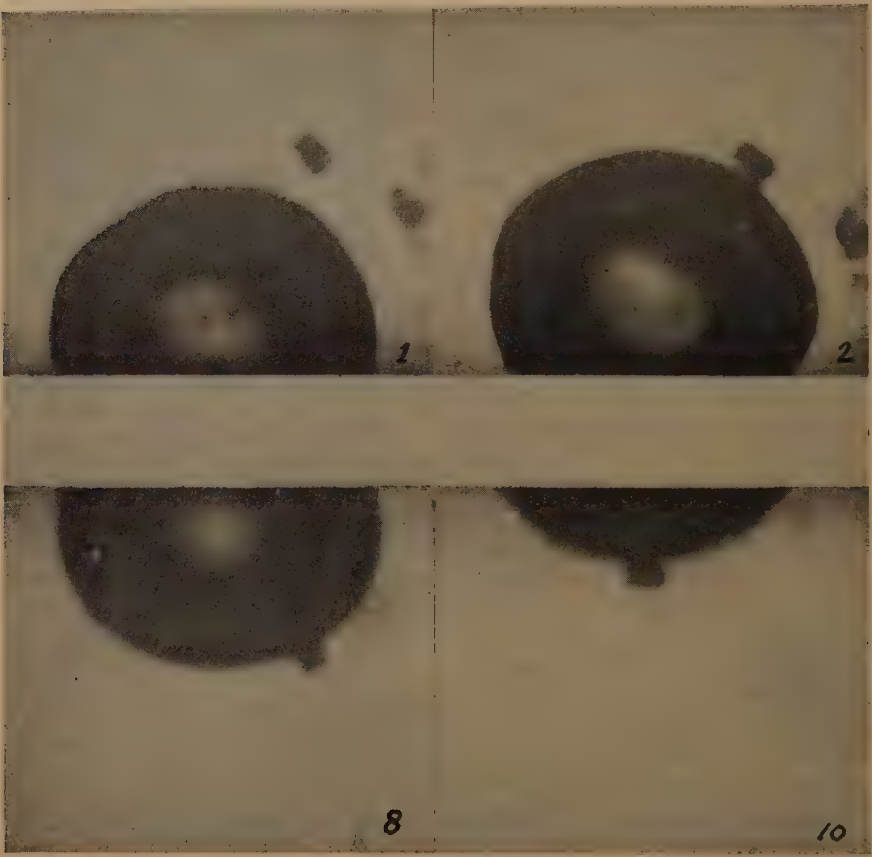

FIG 2—PARTICLE-BUBBLE CONTACT AND ATTACHMENT.
Sequence of action photographed at a speed of approximately 800 frames per second. Sequence proceeds from left to right with frames 1, 2, 8 and 10 shown. The particle makes contact on the upper surface of the bubble, but off-center, and then slides over the surface while revolving. Note that the axis of the particle has turned through 90° between frames 2 to 8.

This observation may also be verified by calculations made on the basis of Reynolds number.[4] The zones of action around a rising bubble which therefore must be considered are: a zone of laminar flow in front of the bubble and a zone of turbulence directly behind it. The zone of turbulence extends roughly to the equator of the bubble. In the photographic record, the laminar flow of liquid to either side of the rising bubble is seen to have a definite effect on the paths of falling particles. The effect is greater on smaller particles, and it is easily seen that very fine particles have but small chance of contacting any individual bubble since fluid-flow lines literally sweep them aside as the bubble rises. This is the gist of Gaudin's theory. Particles in the size range of 100 mesh, however, deviate slightly from a flow-line path when that path is not vertical (and probably also because of their own movement against the fluid). This may explain why they are seen to make contact, off center, on the forward surface of the bubble. The zone of turbulence in the bubble's wake is seen to be another important cause of particle-bubble contact; this phenomenon, which was unpredicted by existing theories is important as it requires an amended theory. Particles which do not make contact on the upper surface of the bubble are drawn many times to the under side of the bubble where contact is effected. Much work of a statistical nature remains to be done before the question can be answered as to the relative importance of contact on the forward surface of the bubble versus contact in the turbulent zone. At present all that can be said is that both occur.

Typical particle-bubble contacts usually result in firm particle-bubble attachment (Fig 2). When such attachment takes place, the particle moves over the bubble surface to the lowest accessible point on the bubble and then rises with the bubble toward the froth zone.

Many contacts do not result in attachment, however, and in fact the particle often actually rebounds an appreciable distance after such a sterile contact. This does not mean that attachment in such a case is impossible because some particles which have thus been rejected on the first contact have become firmly attached to a succeeding bubble.

Limitations in illumination and depth of focus prevented the observation of all contacts on any one bubble; thus only those occurring on the periphery normal to the axis of the microscope could be seen.

Oscillation of Bubble Walls

An interesting and unexpected fact is recorded on the motion micrographs made in this study. The bubbles photographed were produced from a capillary orifice directly below the field of view of the camera. The bubbles produced from this orifice are seen to be oscillating—their walls are in constant and more or less regular motion. The oscillations appear to

FIG 3—BUBBLE IN OSCILLATION.
Sequence proceeds from bottom to top as bubble rises. Note the similarity of frames 1, 4 and 7. Camera speed, 3000 frames per second.

have a frequency of the order of 1000 per second (Fig 3).

This distortion of the bubbles is apparently a result of their release from the orifice, for they cease oscillating and assume a spherical shape after moving a short distance through the solution. The question arises, however, whether or not bubbles are induced to oscillate in this manner by other complex forces such as exist in a violently agitated flotation cell. For example, it seems highly probable that bubbles in the vicinity of a rapidly rotating impeller in a flotation cell may behave in a similar manner.

The effect of bubble oscillation upon attachment of mineral particles to the bubbles is very interesting. Particles in the path of an advancing protuberance of the bubble wall are seen to rebound therefrom. Particles contacting a receding protuberance attach to it quite readily. The overall effect of such oscillations is unknown. The effective diameter of the bubble is obviously increased by the amplitude of the oscillations, but whether or not this is an important factor in flotation remains to be proved.

CONCLUSIONS

Collector-coated mineral particles can and do attach to coursing, free air bubbles. This is not to say that each and every mineral particle which contacts the surface of a bubble attaches to it. Many particles that appear to make contact do not attach. Nevertheless, a large proportion of collector-coated mineral particles which contact a rising, free air bubble attach to the bubble surface.

These experiments were carried out in a very dilute pulp. Each bubble was rising singly through a thin stream of settling mineral particles. Under these conditions, collector-coated mineral particles contact and attach themselves to the rising bubbles. It would seem, therefore, that mineral particles in a thick, swirling pulp containing a large number of bubbles would attach to the bubbles even more readily.

In summary, it has been found that:

1. Collector-coated mineral particles do attach to free air bubbles as a consequence of simple contact between the two.

2. Fine particles are influenced by flow of fluid around the bubble surface to a greater degree than are coarser sizes (but contact of flotation size particles with the bubble is not prevented.)

3. Bubbles released from an orifice are not spherical but exhibit oscillatory motion. It is possible that bubbles may exhibit oscillation when subject to external forces other than release from an orifice.

SUGGESTED RESEARCH

A continued study of the subject of this paper is needed to amplify and clarify the data already secured. We feel that we have only scratched the surface of an interesting and important subject. Our work has been limited to photographic speeds of 3000 frames per second. Much higher speeds are definitely desirable.

The work has also been limited to one type of experimental set up. We can only speculate on the knowledge to be gained by photographing pulp action in the region of an impeller.

Likewise the effects of bubble oscillation demand further study. By the use of high frequency vibrations we may find that particle-bubble contact is improved. Such information would be valuable.

ACKNOWLEDGMENT

The authors wish to acknowledge the inspiration and guidance of Professors A. M. Gaudin and R. Schuhmann, Jr., during the progress of this work.

ILLUSTRATIONS

The movement of a particle in the vicinity of an air bubble when photographed at 3000 frames per second covers a total of 85 frames. Since a single frame or even a few frames cannot convey the proper conception of the motions involved, few photographs are reproduced herewith. A much better idea of particle-bubble contact and attachment may be secured from viewing the original films as projected in motion; therefore copies are available on loan from the Department of Metallurgy, Massachusetts Institute of Technology, Cambridge 39, Mass. The film is 16 mm; running time, 10 min.

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Flotation of Low Grade Gold Ores at Golden Cycle Corporation

By Howard R. Keil*

(New York Meeting, February 1948)

Prior to World War II, approximately 500 tons per day, or one-third of the ore being shipped to the Golden Cycle Mill at Colorado Springs, Colo., from the Cripple Creek district, was being treated in the flotation plant. The remaining 1000 tons, along with the flotation concentrates, received the usual roasting and cyanide treatment. Since it was not possible to make a disposable tailing on the general run of Cripple Creek ores by flotation alone, it was necessary to treat the flotation tailings by cyanidation, along with the roasted ore. By this treatment, an overall economical tailing could be made.

The gold ores from the Cripple Creek district are the highly siliceous sulphotellurides with approximately 3 pct total sulphides. In the lower grade ores most of the values are found in very finely disseminated pyrite. In some of the higher grade ores the tellurides, calaverite and sylvanite, can be recognized.

After several years of floating Cripple Creek gold ores, having a very low mineral content, it was concluded that this type of ore required a type of flotation machine that would give a rather violent aeration and agitation. This action produced a fairly deep froth column which contained not only the free mineral, principally pyrite, but also most of the middling product; this was most essential in producing a low grade tailing by flotation alone that could be discarded to waste. After investigating most of the current types of flotation machines, it was decided to convert to the Fagergren type.

During World War II, the flotation plant was modified to handle lead and zinc ores. The treatment of these ores was carried on until April 1946. At this time the flotation plant was again reconverted to handle the Cripple Creek ores.

Selecting Ores for Flotation

During the period of reconversion a considerable amount of experimental work was carried on in the laboratory to determine if a disposable tailing could be made by flotation alone, on certain selected ores from the Cripple Creek district. It was found that most of the dump ores and a few of the low grade mine ores could be treated by this method with very satisfactory results.

After the preliminary test work in the laboratory, it was decided to carry on further work on a plant-size scale to see if similar results could be obtained. On April 15, 1946, a part of the plant was put into operation as a pilot plant. The various ores that had been found suitable for treatment by this method in the laboratory, were run through the pilot plant at the rate of 200 tons per day for a period of several months. Results were so encouraging that it was decided to put the plant in full-scale operation.

Since that time plant capacity has been increased from 500 to 1000 tons per day. This has been done not by the installation of more machinery but by (1) crush-

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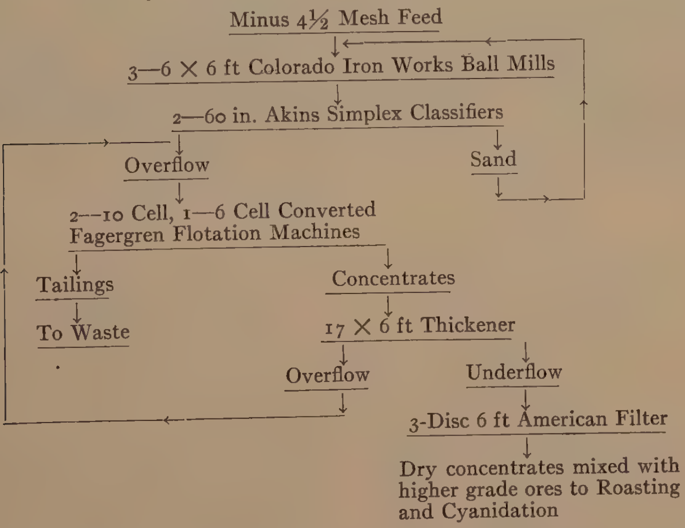
* Assistant Mill Superintendent, Mill Dept., Golden Cycle Corp., Colorado Springs, Colo.

ing the ball-mill feed finer in the secondary crushing plant and (2) grinding coarser in the ball mills. The ball mill–classifier circuit has been the bottle neck as far as plant capacity has been concerned to date. After putting the plant into operation, it was discovered that equal recovery could be made with a much coarser grind than had at first been anticipated, and so by the above mentioned methods it has been possible to double plant capacity without impairing metallurgical results.

FLOWSHEET OF FLOTATION PLANT


Minus 4½ Mesh Feed
3—6 × 6 ft Colorado Iron Works Ball Mills
2—60 in. Akins Simplex Classifiers
Overflow
Sand
2—10 Cell, 1—6 Cell Converted Fagergren Flotation Machines
Tailings
To Waste
Concentrates
17 × 6 ft Thickener
Overflow
Underflow
3-Disc 6 ft American Filter
Dry concentrates mixed with higher grade ores to Roasting and Cyanidation

At present, ball-mill feed is received from the secondary crushing plant at minus 4½ mesh. Grinding is done in three 6 by 6 ft Colorado Iron Works ball mills which have been converted from the overflow to the grate discharge type. The three ball mills are in closed circuit with two 60 in. Akins Simplex classifiers, and are so arranged that each classifier receives the entire discharge from one ball mill plus one-half of the discharge from the third mill. The return sands from the two classifiers are, in turn, split between the three ball mills.

Ni-Hard cast iron liners are used in the ball mills, along with forged steel balls. Ball make-up consists of one-half 2½ in., one-quarter 2 in., and one-quarter 1½ in. balls by weight. Ball consumption is approximately 0.9 lb per ton of ore milled.

The classifier overflow, containing approximately 44 pct solids, is diluted to approximately 30 pct solids and pumped by a 4 in. Wilfley pump to the first of three banks of flotation cells, operating in series; this dilution of pulp reduces the power required in the cells. The first two banks of flotation machines each have ten cells and the third bank has six. The flotation machines, as converted to the Fagergren type, have proved to be very satisfactory on this type of ore. The increased speed and agitation of the Fagergren machines have permitted the treatment of a large tonnage of relatively coarse material with very good metallurgical results. This increased aeration and agitation plays a most important part in making a low-grade tailing as previously mentioned. It has been found that if the top of the froth column is skimmed the tailings will double or treble but by "digging deep" into the froth column—that is, carrying a fairly high pulp level in the cells—and taking off most of the froth, the tailings immediately go down to 0.01 oz gold per ton or lower. This practice is not possible with machines that carry only a shallow froth column. The frother consumption is also considerably less with the converted Fagergren-type machines.

TABLE 1—*Screen Analysis of Ball-mill Feed*

Mesh	Weight, Per Cent	Cummulative Per Cent
Plus 6	5.6	
8	18.5	24.1
10	17.8	41.9
14	11.1	53.0
20	10.1	63.1
28	6.4	69.5
35	5.6	75.1
48	4.1	79.2
65	1.9	81.1
100	4.5	85.6
150	1.5	87.1
200	1.5	88.6
Minus 200	11.4	

TABLE 2—*Screen Analysis of Classifier Overflow*[a]

Mesh	Weight, Per Cent	Cummulative Per Cent
Plus 20	0.6	
28	3.1	3.7
35	8.6	12.3
48	8.5	20.8
65	9.8	30.6
100	19.2	49.8
150	4.8	54.6
200	6.5	61.1
Minus 200	38.9	

[a] Overflow at approximately 44 pct solids. Tons solids overflow approximately 1000 per 24 hr.

CONVERSION TO FAGERGREN-TYPE MACHINES

The conversion to Fagergren machines was accomplished, with a minimum of labor and materials, by utilizing all possible parts from the previous type machines (Fig 1). The agitating mechanisms and baffles from the previous type machines were removed and modified Fagergren rotor-stator assemblies installed. The converted machines each have a triangular shaped stand, elevated 13 in. above the bottom of the cell by three legs. The stator of a 44 in. Fagergren machine, with stand pipe and dispersion apron attached, rests on top of the stand. The nuts on the lower ends of the four rods that hold the stator together rest in four small pockets in the stand, thus preventing the stand from shifting under the stator and also allowing rapid removal of the stator without disturbing the stand. The entire assembly is held in place by three anchor straps, bolted on one end to the stand-pipe and on the other to the two sides and back of the cell. The space between the bottom of the original cell and the bottom of the stator, sands-up immediately when put into operation, thus automatically raising the bottom of the new cell approximately 13 in.

The bearings, bearing housings and shafts from the previous type machines were utilized on the present machines by merely shortening the shaft and modifying the lower end to accommodate the rotor of a 44 in. Fagergren machine.

The flared or tapered section of the froth overflow side of each cell was removed. This side was then extended vertically to a point approximately 4½ in. above the top of the stator to provide a new froth overflow lip. Guides were welded on each end of the overflow lip to permit raising or lowering of froth overflow, by use of metal slats. Two openings, 12 in. high, 24 in. long, one above the other with 4 in. between openings, were cut in each dividing wall between cells. At the discharge end of each bank of cells a box and overflow gate were provided. The gate can be raised when necessary to empty the cells. The gate guides were extended to the top of the discharge box to allow the addition of metal slats for proper control of pulp level and froth removal (Fig 2).

The same 10 hp motors, one motor to each two cells, that were used on the previous type machines, are being used on the converted machines. The rotor speed of

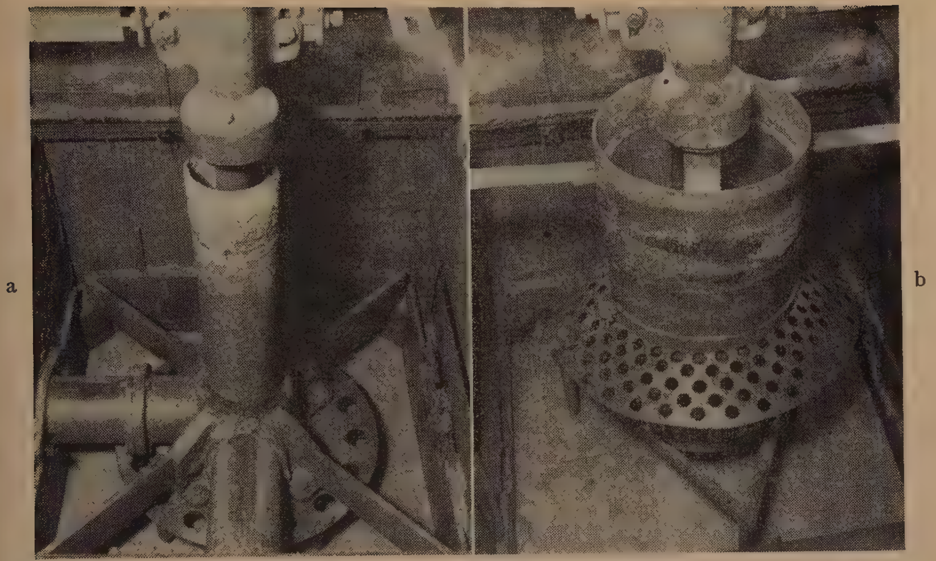


FIG 1—(a) FLOTATION MACHINE BEFORE CONVERSION TO FAGERGREN MACHINE. (b) FLOTATION MACHINE AFTER CONVERSION TO FAGERGREN MACHINE.

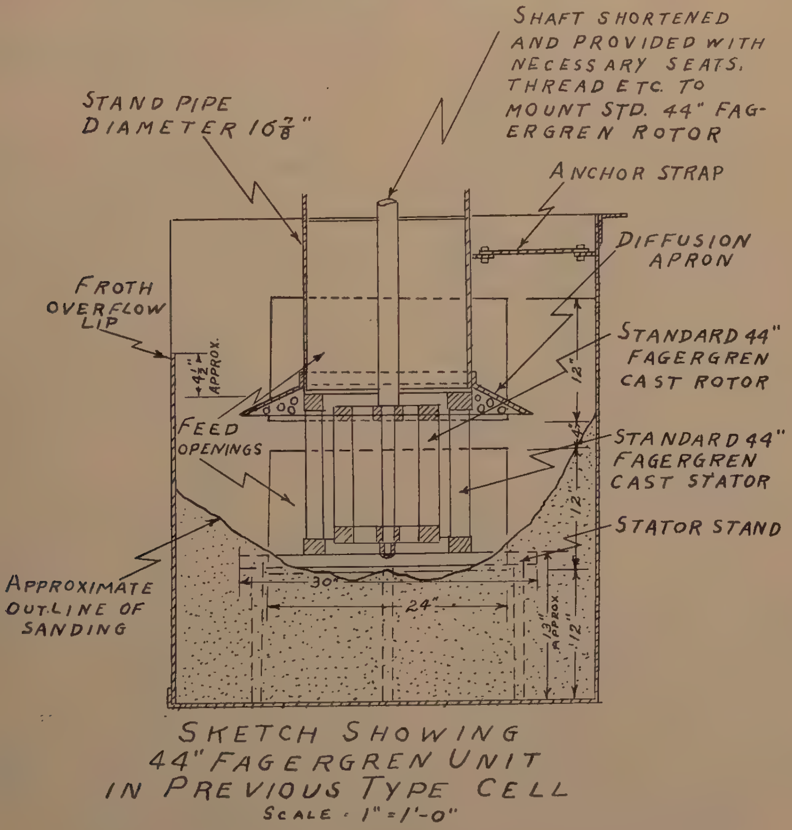

SHAFT SHORTENED AND PROVIDED WITH NECESSARY SEATS, THREAD ETC. TO MOUNT STD. 44" FAGERGREN ROTOR
STAND PIPE DIAMETER 16⅞"
ANCHOR STRAP
DIFFUSION APRON
FROTH OVERFLOW LIP
4½" APPROX.
12"
STANDARD 44" FAGERGREN CAST ROTOR
FEED OPENINGS
4"
STANDARD 44" FAGERGREN CAST STATOR
12"
STATOR STAND
APPROXIMATE OUTLINE OF SANDING
30"
24"
13" APPROX
12"
SKETCH SHOWING 44" FAGERGREN UNIT IN PREVIOUS TYPE CELL
SCALE · 1" = 1'-0"

FIG 2—SKETCH OF 44 IN. FAGERGREN UNIT IN PREVIOUS TYPE CELL.

each cell was increased to plus or minus 640 rpm by changing the sizes of the V-belt drive sheaves. Power required is approximately 9.5 hp for each two cells at the present pulp density.

The rotor and stator bars and spacer rings have been modified somewhat since these machines were converted but are essentially the same as standard Fagergren parts. Spacer rings and bars, as used at present, are made of Ni-Hard cast iron. The life of the rotor is from six to eight weeks and that of the stator from five to six months; this would be increased if it were possible to obtain a better grade of Ni-Hard. A rotor can be changed in approximately 30 min without by-passing or shutting down the entire bank of cells. When replacing stators, however, it is necessary to shut down and drain the cells.

TREATMENT OF CONCENTRATES

The combined rougher concentrates from the three banks of flotation cells are pumped by a 2 in. Wilfley pump to a 17 by 6 ft thickener. The thickened underflow is filtered on a three-disc, 6 ft American filter. Since the thickening area provided by the thickener is not ample to give a clear overflow, it is necessary to feed this overflow back to the head end of the first bank of flotation cells along with the classifier overflow. A larger thickener is under construction at the present time. The flotation reagents used are given in Table 3.

TABLE 3—*Flotation Reagents*

REAGENT	LB. PER TON ORE
American Cyanamid Reagent No. 301, Z4 Xanthate (50–50)	0.20
Pine oil	0.03
Methyl amyl alcohol	0.02

The flotation concentrates, after filtering, run approximately 9 pct moisture, and contain enough coarse middling particles to give a product that breaks up easily, this being quite important in its subsequent treatment. These concentrates are mixed with the higher grade Cripple Creek ores ahead of the roasters and receive the usual roasting and cyanide treatment. Since the concentrates are processed in other sections of the plant in place of being shipped to a smelter, recovery, instead of grade of concentrate, is more important. This is the reason for not cleaning the rougher concentrates. The average assay of gold is given in Table 4.

TABLE 4—*Average Assays Au*

	OZ. PER TON
Heads	0.085
Concentrates	1.50
Tails	0.01
Ratio of concentration	19.8 to 1
Recovery, pct	88.9

TAILINGS

The tailings from the third bank of flotation cells are fed through an 8 in. pipe line, by gravity, to the tailings dam.

CONCLUSION

The converted Fagergren cells have worked very satisfactorily on this type of ore.

Test work is being continually carried on in the laboratory on the various ores from the Cripple Creek district to find out if more of them can be treated by this method. Ways and means are continually being sought to increase plant capacity so that more and more of these low-grade ores can be handled.

ACKNOWLEDGMENTS

The writer wishes to acknowledge the helpful suggestions given him by Marvin F. Dycus, Mill Superintendent, and Max W. Bowen, Mill Manager, in preparation of this article.

Simultaneous Grinding and Flotation

By A. Kenneth Schellinger,* Junior Member and O. Cutler Shepard,* Member AIME

(San Francisco Meeting, February 1949)

Introduction

Overgrinding, or the breaking of ore particles into sizes smaller than required for liberation, is a first-magnitude problem in grinding for concentration processes. The conventional ball mill-classifier circuit used in commercial grinding produces mineral particles that are all ground to pass an upper size limit, while the proportion of very fine sizes, or slimes, is uncontrolled. Calculations based on measurements made in commercial plants which were grinding ore for flotation concentration indicate that about 90 pct of the energy used in grinding is consumed in useless overgrinding.[1]

In addition to accounting for the consumption of a large amount of energy, overgrinding is detrimental to the flotation process. Overground slime coats larger particles and inhibits selective action by the flotation reagents on particles which by themselves could be separated very well. A further difficulty is that extremely finely ground minerals do not respond satisfactorily to flotation-collecting reagents, and both selection and recovery decrease markedly with particles finer than 5 to 10 microns.[2]

Conventional classifier-ball mill circuits inherently tend to overgrind the valuable minerals more than the gangue minerals because most valuable minerals are of higher density and are softer and more friable than the gangue minerals. The remedy to this situation would seem to be a separating mechanism which would separate free valuable mineral particles regardless of size and density considerations and a mechanism which would operate in the grinding mill so that valuable mineral grains would be removed from the grinding machine as soon as they are freed. A consideration of these ideas led the authors to the conception of a simultaneous grinding and flotation mechanism. So far as the authors are aware, such a machine has never been tried, even on a laboratory scale.

Development of Apparatus and Experimental Method

Apparatus development went through several stages before a really workable machine was devised. Work was started with a large cast-iron mortar 7½ in. in diameter and 8 in. deep. A cast-iron disk was mounted on the end of a vertical shaft so that it fitted loosely into the bottom of the mortar. On top of the disk four radial cleats ¼ in. high were bolted to agitate the ball charge as the disk, or impeller, rotated. The vertical shaft was driven at 100 rpm by a mechanism improvised from an old coffee-mill type of grinder.

A hole was drilled into the bottom of the mortar through which compressed air was introduced for frothing. In operation, weighed charges of ore which had been crushed to pass an 8-mesh screen were placed in the bowl along with water and suitable reagents. The impeller was then started and the compressed air turned on so that the ore was ground while being exposed to air bubbles. Mineralized froth was

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[1] References are at the end of the paper.

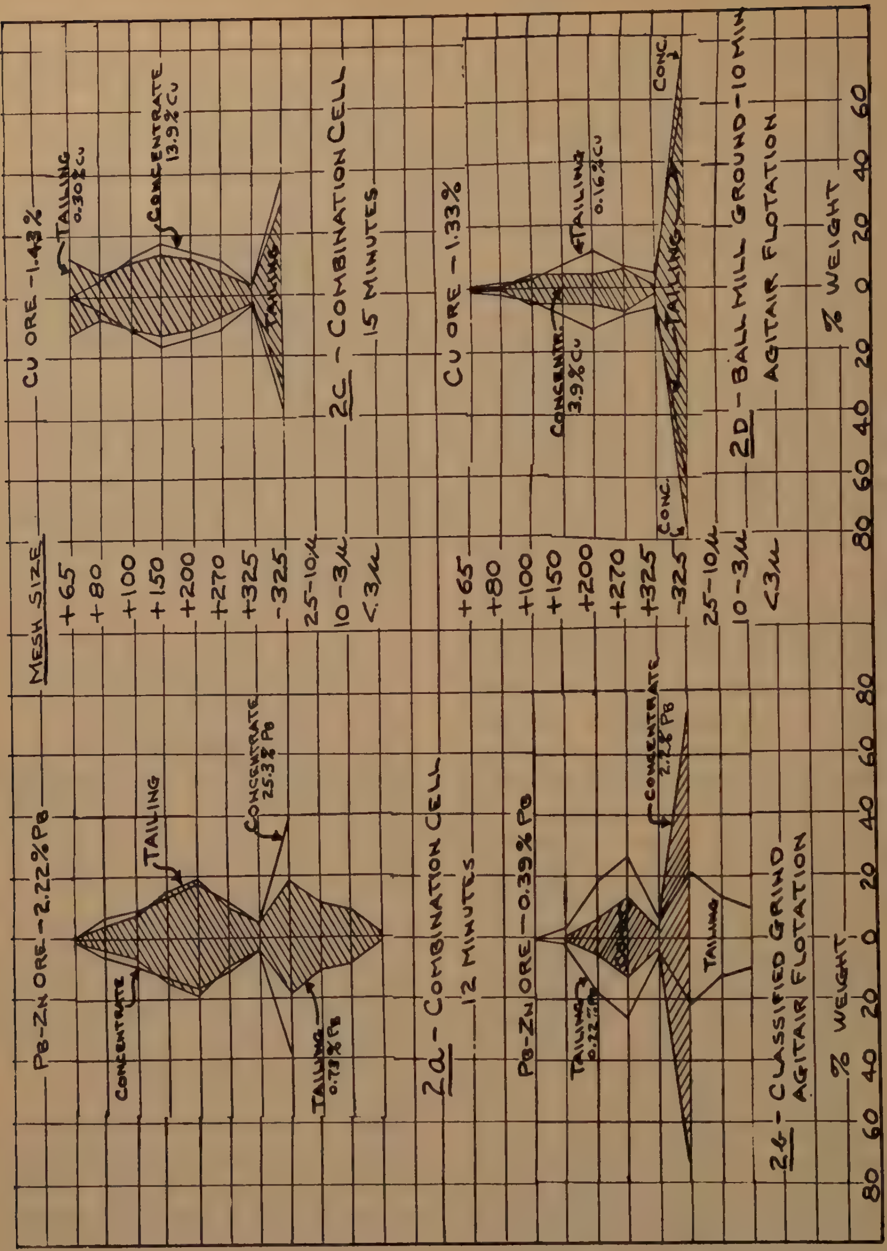

Fig 1—Product size analyses plotted bilaterally.

skimmed from the bowl top into an encircling launder. Excessive pulp swirling suggested the deepening of the pulp column by the addition of a 10-in. pipe section to the top of the bowl. Radial baffle blades were also installed in the pulp column above the balls. These modifications provided a workable apparatus in which a number of batch tests were run on a lead-zinc ore.

An effort was made to compare this method of grinding and flotation with conventional methods. Batches of ore were ground in a laboratory ball mill and treated by flotation in an ordinary laboratory flotation cell. A laboratory ball mill-classifier circuit was also used to grind a large diluted batch of the same ore. A portion of this grind was treated by flotation in a laboratory batch cell. Flotation products from the combination machine and the conventional flotation cell treatments of other grinds were sized by screening and elutriation. Lead analyses were made on suitable samples of the products. The results obtained in the different types of tests are shown in Table 1, and the particle size distributions are graphically illustrated in Fig 1.

TABLE 1—*Comparison of Test Results: Simultaneous Grind-float Apparatus and Conventional Laboratory Apparatus, Lead Ore*

Test Number	1	7	8	9
Apparatus used	Ball Mill Agitair Cell	Grind-float Apparatus	Grind-float Apparatus	Ball Mill-classifier Agitair Cell
Grinding time, minutes	10	12	15	Classifier Overflow
Collector Reagents				
Aerofloat 15, lb per ton	0.10	0.25	0.23	0.23
Xanthate	0.15 K Eth.	0.10 K Pent.	0.10 K Pent.	0.10 K Pent.
Frother Reagents				
Pine oil, lb per ton	0.16	0.24	0.24	0.24
Conditioner Reagents				
Lime, lb per ton	2.0	0	0	0
Na_2CO_3	0 lb per ton	to pH 9	to pH 9	to pH 9
$CuSO_4$, lb per ton	0.80	0	0	0
Head				
Grams weight	572	990.3	1007.5	437
Assay (calc.) Pb, pct	2.24	2.22	2.02	0.39
Pulp density, pct	20	12	12	18
Concentrate				
Grams weight	87	60.3	50.5	37.0
Assay, Pb, pct	12.6	25.3	23.6	2.2
Total Pb, pct	85.3	69.2	58.6	48.5
Minus 325 mesh, weight, pct		38.14	55.0	77.5
Tailing				
Grams weight	485	930	957	400
Assay, Pb, pct	0.39	0.73	0.88	0.22
Total Pb, pct	14.7	30.8	41.4	51.6
Minus 325 mesh, weight, pct	76.5(−200)	39.5	37.4	47.6

NOTE: Ball mill refers to a 12 by 12 in. Braun Laboratory Mill with 40 lb of steel balls.
Agitair cell is a vertical 9 by 5½ by 5½ in. type.
Ball mill-classifier refers to a laboratory demonstration unit consisting of a Denver 14 by 14 in. continuous ball mill in closed circuit with a 48 by 8 in. Dorr Simplex classifier.

Unsatisfactory frothing action brought about the second major revision of the combination machine. The original machine had a tendency to deliver the air in large bubbles at the pulp surface. To break up the air bubbles a small air impeller was placed below the ball impeller. This air impeller was driven at 1500 rpm by a shaft entering through a packing gland in the bottom of the bowl. This impeller sucked in air so that compressed air was not needed. It also dispersed the air into small bubbles which gave better frothing and improved flotation of released sulphide minerals. Fig 2 shows a cross-sectional sketch of the revised combination machine and a photograph is shown in Fig 3.

A series of batch tests was made in the revised apparatus. In these tests a copper ore from the Holden mine was used. Comparison tests were again run in a conventional laboratory batch mill and flotation cell. Results from some typical tests are tabulated in Table 2. Sizing data on concentrates and tailings are summarized graphically in Fig 1.

DISCUSSION OF RESULTS

Enough work on the grind-float process has been done to satisfy the authors that simultaneous grinding and flotation is a feasible process. Laboratory concentration experiments were not carried to great refinement because excellent concentration results in batch laboratory tests would not necessarily mean that continuous commercial machines would be successful. The grade of concentrate in all of the copper-ore tests was low because flotation conditions were not adjusted for the depression of pyrite. The tests do show that flotation can be carried out in a grinding machine and that by this simultaneous grind-float process recovery can be made with less overgrinding than usual. It is believed that by adjusting grinding time and reagents normal recovery and grade of concentrate can be obtained. It is possible that a longer grinding time would be required to increase the recovery of valuable mineral. This would result in grinding the tailing a little finer but could not affect the size of the valuable minerals, since they are removed as fast as they are freed.

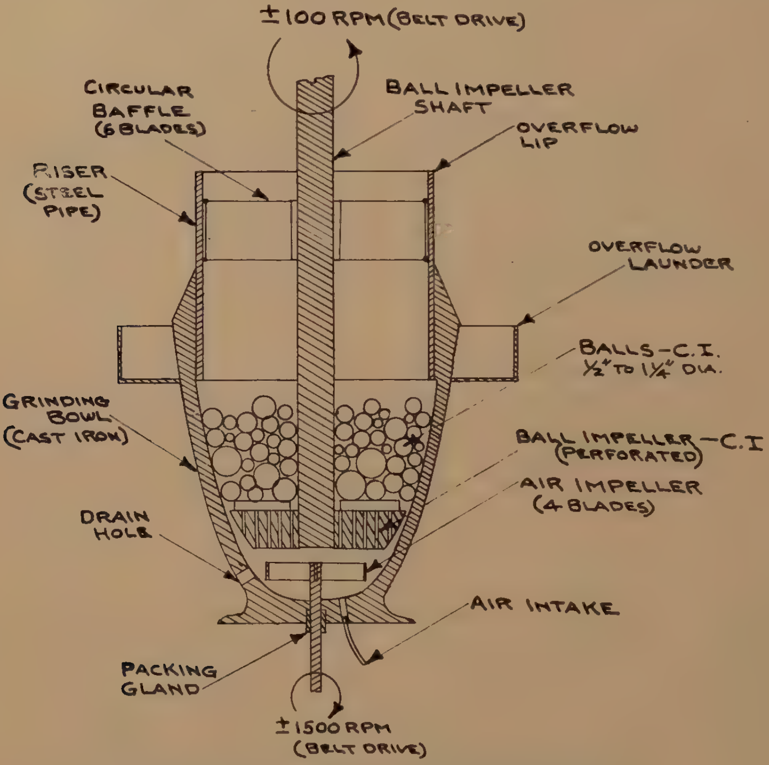

FIG 2—REVISED COMBINATION GRINDING-FLOTATION CELL.
Scale: 1 to 4; reduced approximately one-half.

The authors have considered the possibility of reducing the time in the grind-float machine and then recovering coarse material from the tailing in a conventional classifier for return to the feed of the grind-float machine. However, interlocked grains are not of any definite size and since interlocked minerals usually break more easily than massive minerals it is believed that in most cases regulation of time in the grind-float machine will give good recovery with less useless grinding of gangue minerals.

The grind-float process offers promise of milling economies by:

1. Floating a coarser bulk concentrate thereby saving in reagents, grinding, costs and filtration costs.

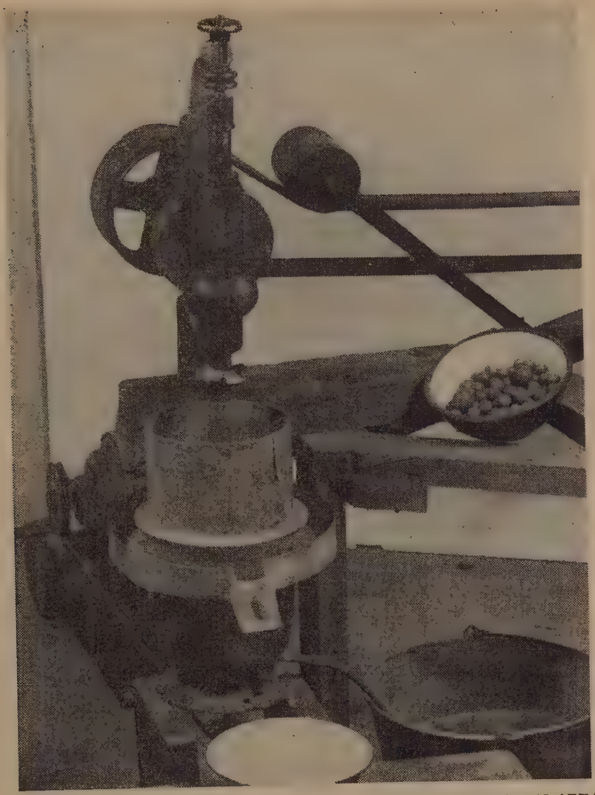

FIG 3—REVISED MODEL OF SIMULTANEOUS GRINDING-FLOTATION APPARATUS.

TABLE 2—*Comparison of Test Results: Revised Simultaneous Grind-float Apparatus and Conventional Laboratory Apparatus, Copper Ore*

Test Number	5A	6A	7A	8A
Apparatus used	Grind-float Apparatus	Ball Mill Agitair Cell	Grind-float Apparatus	Grind-float Apparatus
Grinding time, min	15	10	15	15
Collector Reagents				
Aerofloat 31, lb per ton	0.05	0.05	0.08	0.12
K Pent. Xanthate, lb per ton	0.08	0.08	0.08	0.09
Conditioner Reagents				
Lime	to pH 9	to pH 9	to pH 9	to pH 9
Pine oil, lb per ton	0.06	0.03	0.09	0.15
Head				
Grams weight	1000.1	505	918.5	908
Assay (calc.) Cu, pct	0.71	0.74	0.77	0.70
Concentrate				
Grams weight	30.1	79.0	67.5	100.0
Assay, Cu, pct	13.9	3.9	7.8	5.0
Total Cu, pct	59	81.9	74.7	79.0
Minus 325 mesh, weight, pct	39.4	76.4	65.2	
Tailing				
Grams weight	970	426	851	808
Assay, Cu, pct	0.30	0.160	0.21	0.165
Total Cu, pct	41.0	18.1	25.3	21.0
Minus 325 mesh, weight, pct	33.9	61.3	35.1	41.7

NOTE: See note under Table 1 for description of apparatus.
Revised grind-float apparatus refers to modification of cell by placing a high-speed air impeller under the ball impeller.
Grinding media: 17 lb of steel balls (½ to 1¼ in.) was used in the grind-float apparatus, or 10 lb of flint pebbles (1½ in.). The Braun Laboratory ball mill contained 40 lb of steel balls (½ to ¼ in.).

2. Facilitating discard of a commercial tailing at a coarser size with attendant saving in grinding cost.

3. Probable saving in space and equipment cost because of simplification of operations in the combination machine.

4. Possible increased recovery due to exposure of fresh, valuable mineral surfaces to collecting reagents and due to avoidance of overgrinding valuable minerals.

It is believed that a commercial grind-float machine can be built which will operate satisfactorily. Whether or not such a machine will prove economically successful remains to be demonstrated. It is hoped that this paper will stimulate thought and discussion which will lead to continuing work on a larger scale.

SUMMARY

1. To avoid overgrinding with attendant waste of power and detriment to the flotation separation, a unique technique involving the simultaneous actions of grinding and flotation is proposed.

2. The development of a successful laboratory grind-float machine for experimentation with simultaneous grinding and flotation is described.

3. Results of laboratory grind-float and conventional laboratory tests on two different ores are tabulated and a graphical presentation of the particle size distribution in the products of the different tests is given to show that less overgrinding occurs in the new process.

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2. A. M. Gaudin: Principles of Mineral Dressing. 402. (1939) New York. McGraw-Hill Book Co.

Notes on Drying for Electrostatic Separation of Particles

By Foster Fraas,* Member AIME

(New York Meeting, February 1948)

Introduction

That variations in the humidity of the air and in the moisture content of a mixture of broken solids being separated electrostatically cause trouble is not new.[1] Much of the reputation for unreliability commonly ascribed to the electrostatic separation processes is probably due to this one factor alone. Some operators dry thoroughly before separating; and while this is good in many instances, there are others in which it is not the remedy for trouble. Some mixtures behave best in very dry air and with the mixtures fed to the separator while hot to reduce the adsorbed water content. On the other hand, an overdried mixture on coming to equilibrium with more humid air, has been known to take up moisture selectively on one family of particles, as of coal in the presence of slate and clay, and give sharper electrostatic separation.

From these observations it is concluded that there is a permissible moisture content for a particle and also for the atmosphere surrounding it, if in equilibrium, at which the particle will have a conductivity or contact potential[2–4] that is ideal for electrostatic separation. This permissible moisture content varies between a wide range of limits for the different materials that are separated and with the kind of electrostatic separating process used. Water can be present as a result of adsorption and condensation from the atmosphere, as an accumulation before or during mining, or as a residue from a previous process step. The water may exist as a liquid film or as an adsorbed layer of vapor.

If the water exceeds permissible limits it must be removed. Its state determines the method. Drying by evaporation is necessary for adsorbed water, for part of the liquid-film water (that is, up to several percent of the dry weight), and for free water in the internal structure. Large water contents may be removed by drainage or filtration. At just what point mechanical removal should cease and drying by evaporation begin depends upon the method of mechanical removal and its cost.

Particles can also be separated according to their differences in dielectric constant when suspended in a liquid of intermediate dielectric constant. Recovery of this liquid presents a similar problem.

Permissible Moisture

The adsorbed moisture on the outer surface of minerals can be quickly removed or replaced. Accordingly, in speaking of adsorbed moisture contents, either that of the solid or that of the surrounding air may be considered, the two values being connected by a distribution coefficient which varies with the temperature. However, such a condition does not necessarily exist in processes where an initially dry material is suddenly introduced into an atmosphere having a partial pressure of water vapor or where a moist material is suddenly introduced into a dry atmosphere. Here the resultant differences in the

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[1] References are at the end of the paper.

electrical properties of the constituents may depend on a rate of adsorption, or desorption, rather than on an equilibrium distribution.

Specific examples show that moisture influences the contact potential and consequently the frictional electrification. With quartz and feldspar, without previous treatment except for dedusting, relative humidities of the ambient atmosphere as high as 20 pct are permissible. This has been described in a previous paper by the author.[5] On frictional charging by passage over a vibrating aluminum plate "electrifier," dry particles of quartz and of feldspar are charged negatively in air that is thoroughly dried by silica gel. After the humidity is raised slightly the feldspar becomes positively charged with the quartz still negative. Quartz can be separated from feldspar under these conditions. Subsequent gradual increase of the humidity results in the quartz becoming positively charged, and finally with higher humidities, frictional electrification of both the quartz and feldspar cannot be effected.

In contrast, the permissible moisture content for some of the other minerals is much higher. For example, fluorite and sphalerite can be separated by frictional electrification even under relatively high humidity conditions. The effect of adsorbed moisture in preventing electrification may be due either to a high surface conductivity, with subsequent leakage of the frictional electrical charges as fast as they are formed, or to a similarity of electrical surface properties, a result of the same moisture film adsorbed on all the surfaces.

The permissible moisture content also varies in electrostatic separations that utilize differences in conductance. In separating good conductors, such as chromite and magnetite, from quartz, a very low moisture content is necessary and is usually obtained by heating the mineral combination before separation. If clean zircon is substituted for the quartz, the permissible content of moisture can be much higher, and in many cases heating is not required. The differences in the permissible contents of moisture for quartz and zircon are illustrated in the separation of these minerals from good conductors. When heated, the zircon and quartz are separable as nonconductors from such conductors as chromite. When cold, and with appreciable humidity, the quartz acts as a conductor and can be separated from the zircon as a nonconductor. Obviously, quartz attracts moisture which makes its surface conductive.

Conditioning with reagents before separation frequently decreases sensitivity to high-humidity atmospheres; this has been the apparent effect in the use of benzoic acid as a conditioning reagent.[6]

For separations based on a difference of shape factor, the essential feature is a maximum surface charge and not a difference in the conductivities or frictional chargabilities of the various minerals. An example is presented in the separation of muscovite mica from sand,[7] where the highest humidity (the condition for highest surface conductivity) was found to be most effective.

By placing electrodes at the ends of thin slabs of different materials, Field[8] has measured the surface conductivity of mica, glass, quartz, polysterene, hydrocarbons and other substances in contact with air of various humidities. In 100 pct relative-humidity air, mica, glass and quartz have appreciable surface conductivity, whereas polysterene and certain hydrocarbons remain relatively low. Glass coated with the silicones is also low. The silicones and their derivatives are an interesting series of new carbon-silicon compounds recently described by Rochow.[9] The surface conductivity varies with time. For quartz placed in a 100 pct-humidity atmosphere it increases in 10 minutes, 10^6-fold over that at zero humidity. After this time the conductivity remains constant. The conductivity initially overshoots the final value by ten fold. The return to the lower equilibrium value is explained by the sweeping of the excess ions to the electrodes by the steady potential used in measuring the conductivity. When the potential is temporarily removed, the higher value is again obtained, but the lower equilibrium value is again restored in a few minutes after the potential is reapplied.

Cellulose acetate butyrate, although absorbing as much as 4 pct internal moisture, was one of the lowest conductivity substances measured.

The surface conductivities of solids vary with the relative humidities. Field checks the conclusion of Curtis[10] who found that, under conditions of equilibrium, the conductivity is a logarithmic function of the relative humidity. This may be represented by the equation:

$$\log C = aH + b,$$

where C is the electrical conductivity, H the relative humidity, and a and b constants for any one material but variables for different materials.

In general low-volume conductivity materials having water repellent surfaces will have low-surface conductivities in high-humidity atmospheres. The converse of this is not generally true, as is illustrated with cellulose acetate butyrate, which although absorbing as much as 4 pct internal moisture still has a low conductivity. Trace impurities, capable of ionization, probably are effective in increasing the conductivity under high-humidity conditions. For example, traces of the alkali elements in quartz have the ability to ionize in a water film on the quartz surface to yield appreciable conductivity.

Johnson states[11] that electrostatic concentrates may contain as high as 0.75 to 3.5 pct moisture.

All this information indicates that, for certain substances, the permissible moisture content can be high and that moisture removal need not always be very rigorous.

Centrifuge

The common methods for water removal or dewatering are of the drainage type, typified by the rake classifier, and the air pressure or suction-filtration type. With the smaller particle size ranges, filtration will yield a quicker, more efficient water removal than drainage alone. However, with the coarser sizes, air-pressure filtration has little value. After an air channel once forms in the large size interstices the subsequent removal of water is largely a gravity drainage.

An excellent but so far little used method for dewatering in mineral dressing is centrifuging. Centrifuging has its value in increasing the drainage force over that obtained by gravity alone. Taggart[12] has described the Carpenter continuous-centrifugal filter, consisting of a conical chamber rotating about its vertical conical axis. Liquid is removed by filtration through a conical step screen.

The rotor has a speed of 7500 fpm at its maximum rotor diameter of 78 in. With a fine bituminous slack of less than $\frac{5}{16}$-in. diameter, the water in a feed carrying 30 pct moisture is reduced to a value less than 7 pct. The calculated centripetal acceleration at the maximum diameter is 150 G, or 150 times that of gravity.

A continuous solid-liquid centrifugal separator having a horizontal rotating axis has been recently introduced on the market. Instead of the liquid filtering through the solid, as in the Carpenter type, the solid is dragged out of the liquid in a direction toward the apex end of the rotating truncated cone by means of a spiral rake. The centripetal acceleration is approximately 2000 G. There may be some tendency toward abrasion at such a high acceleration. However, the use of abrasion-resistant materials for construction should solve this difficulty.

On coarse material and with sufficient time for the centrifuging action, all but traces of liquid can be removed. An illustrative example is the dewatering of a minus 14-plus 65-mesh sillimanite—quartz rock in a small, batch-type centrifuge. With heap drainage aided by tamping, the retained water amounted to 20.7 pct, whereas with centrifuge dewatering it was only 0.17 pct. For the finer sizes the efficiency of liquid removal decreases so that at minus 100-mesh there is little advantage over filtration. This is illustrated in Table 1.

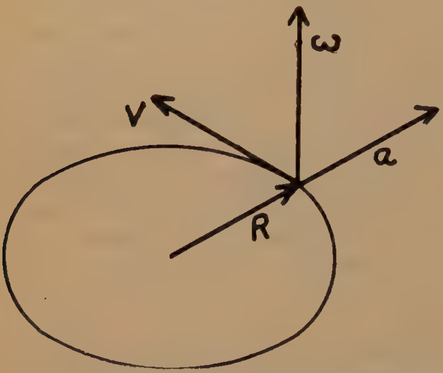

FIG 1—VARIABLES INVOLVED IN CENTRIPETAL ACCELERATION.

TABLE 1—*Dewatering of Minus* 100-*mesh Calcite*[1]

METHOD	RESIDUAL WATER PERCENTAGE OF DRY WEIGHT
Buchner funnel, material not filmed....	9.1
Centrifuge, material not filmed.........	7.2
Centrifuge, material filmed water repellant[2].........................	2.9

[1] Each dewatering was for period of 5 min.

[2] Filmed with 2 lb oleic acid and 0.5 lb Emulsol X1 per ton.

With the centrifuge, minus 100-mesh dewatered calcite still contains 7.2 pct water. However, this is lower than the 9.1 pct obtained by suction filtration. The forces that retain the water are either static, that is, capillary attraction, or dynamic, that is, resistance to fluid flow through small interstices.

Representing by vectors, as shown in Fig 1, the centripetal acceleration, a, the angular velocity, ω, and the tangential velocity, V, all at right angles to each other, the centripetal acceleration equals the vector product:

$$a = V \times \omega$$

The tangential velocity can be expressed in terms of the radius of rotation, R, by the vector product:

$$V = \omega \times R.$$

Eliminating V, we have:

$$a = \omega \times R \times \omega.$$

For magnitude only, the scalar equation for the centripetal acceleration is, accordingly:

$$a = \omega^2 R.$$

With a hypothetical cylindrical pore of radius, r, the capillary attraction is represented by:

$$f = 2\pi r \sigma \cos \theta,$$

where θ is the liquid-solid contact angle, and σ is the surface tension of the liquid. The force resulting from the centripetal acceleration is:

$$f = \pi r^2 h(\rho_l - \rho_g)a,$$

where h is the length of the cylindrical column, a is the acceleration and ρ_l and ρ_g are the densities of the liquid and gas (air). With water this difference approximately equals 1. More exactly, r is the radius of the crown of the meniscus.

The equilibrium is then represented by:

$$2\sigma \cos \theta = rha.$$

With a hypothetical crack instead of a cylindrical pore, where the length of the crack is much greater than the width, r, the same equation is obtained.

In a granular mass the water-filled voids have no resemblance to cylindrical columns.

The voids, however, may be mapped out into irregularly shaped columnar regions oriented with the columnar length parallel, and in some instances at an angle, to the centripetal acceleration. In this case r is not constant but varies continuously. In general the columns will have expansions and constrictions along their length. Except for a variation with respect to the distance from the center of rotation, the force resulting from the centripetal acceleration will be uniformly impressed on all the masses of water. In contrast to this, the opposing capillary force, when liquid is displaced by gas, will not be uniform but will vary with a maximum at the constrictions and a minimum at the expansions. Initially, instead of a capillary force, a resistance to fluid flow through the constriction will be present.

The resulting forces on the water masses will then be a combination of compressions and tensions. When the tension is greater than the atmospheric pressure minus the vapor pressure of the liquid, the columns will break, particularly at the expansions. The columns will also separate because air penetrates from the exterior. The water subsequently flowing from the mass will or will not remain at the constrictions, depending upon whether the radius of the constriction is greater or less than the critical.

Exact calculation of the critical radius depends on the shape of the constriction. The forces resulting from the centripetal acceleration and the capillary attraction will be in equilibrium, with h approaching a value less than r. Under these conditions, the liquid plug tends to puncture and coalesce onto the sides of the constriction. The value of h will probably range from 0.1 r to r. We, therefore, have:

$$h = br,$$

where b may have some value between 0.1 and 1.

The equilibrium is represented by:

$$2\sigma \cos \theta = br^2a.$$

With a tenfold change in b, it would seem needless to have accurate values for σ and θ. This, however, is not generally true, since, by using the same particle size and shape, which is equivalent to holding b constant, comparisons can be made between different values for σ and θ.

Using calcite and water, $\sigma = 72.8$, $\theta = 70°$ and $a = 2000$ G, the calculated value of r ranges from 50 to 150 microns. G is the acceleration of gravity, which in the cgs system has the value of 980. For pore spaces of radii less than the critical, the water will remain, and for those of radii greater, the water will be removed, by centrifuging. The water will also remain in pore spaces having greater than the critical radius if the orientation is at an angle with the centripetal acceleration.

The equilibrium conditions just described represent the maximum volume of water that can be removed with a sufficient time for centrifuging. For shorter periods, the volume of water removed depends on the fluid or water flow factors.

By filming the solid so that it is water repellent, the capillary attraction is eliminated. Table 1 shows that without capillary attraction the residual water content of the above calcite is reduced from 7.2 to 2.9 pct. The remaining 2.9 pct may be present as a result of either an insufficient time for fluid flow or a mechanical obstruction to droplets.

The batch type centrifuge used in all these tests had a perforated basket with a 4.9 in. id.* The perforations were covered with filter paper. With the measured speed of 4900 to 5200 rpm, the calculated centripetal acceleration is 1760 G.

In the continuous horizontal-axis types, the solid is removed from the liquid by spiral rakes. The churning of the solid by the rakes after it is removed from the liquid may be beneficial to further reduction in residual water. Any churning or mechanical movement during application of the centripetal acceleration would release mechanically obstructed droplets and also open constrictions with less than the critical radius.

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* International Equipment Co., No. P6561.

Although the difference in the residual water with and without filming is not large, *the difference in appearance is pronounced.* The unfilmed calcite is aggregated into lumps, whereas the filmed calcite is almost free flowing.

SEPARATION AFTER CENTRIFUGING

Since the centrifuge can reduce the moisture content of coarsely sized material to very low values, a value below the permissible moisture for separation may possibly be reached for some materials. Materials to be selected for examples could be either those which have natural high permissible moisture values or those which may be coated with films having these high values. The films may also be water repellent and aid in the dewatering.

The effect of filming with oleic acid and a hydrocarbon oil is illustrated with a siliceous fluorite from Rosiclare, Illinois. A minus 10-plus 35-mesh sample was deslimed with water by agitation and then treated in a 65 pct solids pulp with 2 lb per ton of sodium silicate. After several minutes an emulsion of 0.6 lb oleic acid, 0.5 lb of Emulsol X1, and 1 lb per ton of transformer oil was added and the agitation continued for 5 min. The slurry was then dewatered in the above described centrifuge and immediately transferred to the electrostatic separator, which was not heated. The centrifuged solids were free-flowing, except for the finest size close to 35 mesh, where enough liquid was present to result in some adhesion between the particles. The results are summarized in Table 2.

The separator was of the simple roll type with an aluminum vibrating feed plate-electrifier[5] shown as item 1 of Fig 1 in a previous publication.[13] It is a unit suited to either conductance or frictional electric separations. A single-roll unit was used but the material was repassed to simulate a commercial multiroll separator. The flow sheet was as follows: the heading was initially passed through a 6-roll separator, with the deflected fraction or deflectant[13] passing for re-treatment to succeeding rolls. The combined undeflected fractions were then passed through a 3-roll separator, and the undeflected fractions of each re-treated on succeeding rolls. The final deflectant from the 6-roll separator constituted fraction 1 or the silica tailing, the final residue from the 3-roll separator, the fluorite concentrate or fraction 3, and the combined deflectants from the 3-roll separator, the middling or fraction 2.

TABLE 2—*Concentration of Siliceous Fluorite after Centrifuging*[1]

Product		Weight, Pct	Assay, Pct		Percentage of Total Fluorite
No.	Class		Silica[2]	Fluorite	
1	Silica tailing....	8.0	73.9	26.1	2.6
2	Middling.......	26.1	45.1	54.9	17.7
3	Fluorite concentrate..........	65.9	1.6	98.4	79.7
	Composite.....	100.0	18.6	81.4	100.0

[1] Minus 10-plus 35-mesh.

[2] As residue by dissolving fluorite in aqueous 25 pct HCl containing 2 pct H_3BO_3.

The electrodes were of positive polarity. For the definition of the electrode a previous paper on the conductance-electrostatic separator may be consulted.[13] Since approximately the same results are obtained with a negative electrode, the separation apparently depends only on conductivity differences. Under these conditions the vibrating plate can be omitted as a frictional electrifier. It functioned merely as a feeder. Non-conducting oils, other than transformer oil, or even fusible insulating solids could be used. Without reagents and with only centrifuging of water from the deslimed solid, no electrostatic separation is obtained.

The same procedure and reagents were applied to a kyanite-quartz, and a synthetic calcite-quartz mixture. However, contrary to expectations, both the oleic-acid-and-oil-coated kyanite and calcite were deflected while the quartz was not deflected. Several other factors in addition to the water-repellent characteristics of an oleic acid coated surface may be influential. One of these is surface texture. The fluorite had the smooth cleavage faces of transparent (macroscopic) crystals, whereas its siliceous gangue, petrographically identified as quartz, was rough and opaque (macroscopic). In contrast to this, the quartz associated with the kyanite and calcite was transparent with smooth surfaces. Surface pore spaces and cracks on a rough, opaque mineral surface may retain enough water for electrical conduction and consequently defeat the nonconductive characteristics of the oleic-acid film. The lesson to be drawn in these experiments is that, whereas certain conditioning treatments to alter surfaces often bring predictable results, it is well to test every sample thought likely to yield the desired results. Reverse effects from predictions do take place and must be watched for.

A similar behavior was found with a feldspar-quartz mixture, the feldspar being coated with oleic acid and oil, using hydrofluoric acid as an activator. These factors probably should be made to operate in unison, such as by coating the quartz instead of the feldspar. Efficient separation was obtained by merely treating the feldspar-quartz in an aqueous pulp with 2 lb per ton of hydrofluoric acid. After dewatering in a centrifuge and passage through a cold electrostatic separator the results of Table 3 were obtained. No other drying was used before electrostatic separation.

The electrostatic separation utilized only three rolls. The separator had the aluminum feeder plate-electrifier, previously described for the fluorite separation, and a negative polarity electrode. The first roll made a separation into a rough feldspar concentrate and a rough quartz concentrate. Passage of each of the rough concentrates over another roll gave the recleaned concentrates of Table 3. The quartz concentrate (fraction 4) assayed 96 pct quartz. The assay of the feldspar concentrate (86 pct) can be raised and the reclean tailings recovered by further passage. With a positive electrode there was not much separation. This would indicate that charging by contact potential on the vibrating plate-electrifier is an important factor for this particular separation.

TABLE 3—*Separation of Quartz-feldspar after Centrifuging*[1]

Product		Weight, Pct	Feldspar Assay, Pct	Percentage of Total	
No.	Class			Feldspar	Quartz
1	Deflectant of 1st deflectant.....	23.0	86	62.8	4.7
2	Residue of 1st deflectant.....	20.9	22	14.6	23.8
3	Deflectant of 1st residue........	6.0	85	16.2	1.3
4	Residue of 1st residue........	50.1	4	6.4	70.2
	Composite.....	100.0	32	100.0	100.0

[1] Minus 10 plus 35-mesh.

[2] By handpicking and weighing. Remainder is quartz.

The test was conducted on the only feldspar-quartz sample available. Since the sample was partly kaolinized, the kaolin part was first removed by aqueous scrubbing and desliming. The method of scrubbing was that described by Norman and Ralston.[14]

Without any hydrofluoric acid treatment, no electrostatic separation could be obtained after centrifugal dewatering with either a positive or negative electrode.

Nothing is known concerning drying by the ambient air during these separations. If drying did occur it was rapid, since the separation time was brief. However, drying by the ambient air under high-humidity conditions could be assured by heating the separator. A modified process for the quartz-feldspar example is the complete drying with heat of the rough feldspar concentrate as a preliminary step to the recleaning operation.

In his paper on the economics of electrostatic separation, Bullock[15] states that electrostatic separation can be obtained on feed as coarse as minus 6-mesh. These sizes agree with the standard screen scale, corresponding, for 6-, 10- and 35-mesh, to openings of 3.36, 2 and 0.5 millimeters. In general, approximately 75 pct of most minus 10-mesh stage ground feeds will be present in the plus 35-mesh fraction. The minus 35-mesh fraction can be separated electrostatically after drying by evaporation, or it may be separated by froth flotation. If separated electrostatically, the increased cost of drying the minus 35-mesh fraction would be compensated by eliminating the collector reagents required for froth flotation.

Fluorine forms covalent compounds with many elements. Covalent compounds have little ionization. The formation of nonionized compounds with some of the ionizing trace impurities on the quartz surface may explain why fluorine is effective in improving the separation of quartz under high-humidity conditions.

DRYING BY EVAPORATION

Finch[12] states that operating costs in drying are extremely variable because of a large number of influential factors. These include differences in relative humidities, altitude, the construction of the drier and the properties of the material to be dried.

It is, therefore, difficult to reach any conclusions as to the effect of moisture content and particle size on the heat consumption by examining the data of actual operation. However, on a theoretical basis all the factors, except the one to be studied, can be held constant and the effect of the varied factor on the heat consumption determined.

The heat consumption is divided into useful heat and lost heat. Useful heat is that required for evaporating water. Additional heat is necessary for the sensible heat of the material and the water, unless the process is carried out adiabatically, such as by drying in low-humidity air. Lost heat includes that lost from the outer walls of the drier and that which escapes in the heated air or fuel gases not saturated with water vapor. In electrostatic separation the sensible heat in the dried solid can be considered useful for later stages of the flow sheet.

In a nonadiabatic-drying process the total useful heat may be represented by:

Total heat = (sensible heat of material) + (sensible heat of water) + (latent heat of evaporation of water).

With an increase of 100°C in drying, the total heat versus the percentage of moisture is shown in Fig 2. An arbitrary value of 0.2 has been selected for the specific heat of the material. This is an average value for many minerals. For other values of specific heat the plotted lines would be parallel with the zero moisture intercept at the corresponding specific heat of the solid. That the data of Fig 2 are on the basis of a 100°C rise in temperature of the drying feed, must be emphasized. Calories per gram may be converted to Btu per ton by multiplying by 3600.

Since the latent heat of evaporation is relatively of much greater magnitude than the specific heat, the total heat required increases rapidly with the increase in the percentage of moisture. Thus with a 0.2 specific heat for the solid, the heat requirement for complete drying of a 5.1-percentage of moisture feed is twice that for 1 percentage of moisture feed. For drying a 3.1-percentage of moisture feed, it is twice that required for heating the dry solid.

This demonstrates that drying a feed with an initial low-moisture content will result in an appreciable saving in fuel consumption over that required for a feed having a high-moisture content. This, of course, is under the condition that all other factors, such as plant location, type of drier, and atmospheric conditions, are kept constant.

Another influential factor is particle size. Fine materials that are to be dried are usually caked into lumps or slabs. Removing moisture from the internal regions of such aggregates requires time for diffusion and also for heat transfer. This increased time for moisture and heat transfer for caked fine feeds over that for the simple surface evaporation of moisture from coarse, free-flowing solids results in an increase in the size of drying equipment, with consequent increase in heat losses. For a given method of dewatering, a coarse feed will also have a lower-moisture content than a fine feed. Accordingly there are lower heat losses and a lower initial investment for drying coarse electrostatic feeds than for drying fine froth-flotation concentrates. The use of electrostatic separation on coarse particles to avoid drying after the quartz-feldspar flotation separation of finely ground material is described by Gillson.[4]

A frequent statement defending wet methods of mineral dressing and objecting to dry methods is that the crushed rock as mined is usually wet. However, with a more thorough consideration of the problem it is found that the actual moisture retained by coal or rock is usually small.

If we assume that each particle has a uniform outer layer of water of thickness A, then for a particle of diameter y, fraction X of the dry weight as water is:

$$\frac{B(y+2A)^3}{By^3} = 1 + X.$$

Here B is a shape factor, 1 for a cube, $\frac{\pi}{6}$ for a sphere, etc. This equation is difficult for calculations or to represent graphically, but by taking logarithms to the base e of both sides and expanding these logarithmic functions into series form, a more simplified expression can be obtained.

Noting that $\frac{A}{y}$ and X are small and that terms with powers greater than 1 can be dropped, a good approximation is represented by:

$$\frac{6A}{y} = X$$

If for two different sizes, 1 and 2, we assume the same thickness of water film, A can be eliminated and we have,

$$y_1X_1 = y_2X_2$$

A sample of minus 6 plus 10-mesh quartz-feldspar submerged in water and then drained, without raking, for 2 min on a screen, contained 20 pct of its dry weight as water. By calculation from the last equation, a 7-mesh (2.8-mm) particle retaining 20 pct of its dry weight as water would have a water film corresponding to 1 pct for a 2.3-in.-diameter average size. Accordingly, rock as mined at an average 2.3-in. size, even if drenched, would only retain less than 2 pct surface moisture. For porous and friable materials, however, this relationship does not hold. In this case, moisture is retained not only on the surface but in appreciable quantities in the interior.

COSTS FOR DRYING AND CENTRIFUGING

For the spiral-rake, horizontal-axis-type centrifuge (Bird) recently marketed, the cost depends primarily on the power. A conservative estimate for a 25 ton per hour machine is as follows:

Annual fixed costs

Depreciation, with first cost of $22,000, 10 pct	$2,200	
Interest, 6 pct	1,320	
Insurance, taxes and social security, 2½ pct	550	
Total		$ 4,070

Annual variable costs

Labor, ¼ of total time, 2,190 hours at 75 cents per hour	$1,643	
Power, 75 kw, 8760 hr at 1 cent per kw-hr	6,570	
Upkeep, 1½ pct	330	
Total		8,543
Total annual cost		$12,613
Total annual tonnage		219,000
Cost per ton		5.8 cents

Taggart[12] quotes a cost of less than 2 cents per ton for drying coal in the Carpenter centrifugal filter.

For drying stone from an initial 4.8 to a final 0.3 percentage of moisture in a single shell-type, direct-heat rotary drier at the rate of 21 tons per hour, Sherwood[4] gives a charge of 16 cents per ton. Finch[12] quotes expenses averaging approximately 25 cents per ton but ranging from 10.3 to 60.9 cents. Johnson[11] in his cost estimates for electrostatic separation quotes 25 to 40 cents per ton for drying. He states that if bone dryness is required, the increased cost will be only 1½ to 2 cents per ton.

All these costs are prewar values and must be adjusted to fit the present economic conditions.

If fuel costs are high and power costs low, it may be advantageous to dewater in a centrifuge before drying by evaporation methods. Costs for other dewatering devices, such as rake classifiers, would be eliminated. The centrifuge can also be driven by a variety of prime movers. With a steam prime mover driving the centrifuge, the exhaust steam can be used for drying purposes and for heating the electrostatic separator, such as by passage through the hollow rotating rolls.

A simple, high-speed, atmospheric exhaust steam engine[4] consumes approximately 25 lb of steam per indicated horsepower hour when the pressure supplied to the engine is 175 psi. For a 100 hp drive, equivalent to 75 kw., the steam consumption would be 2500 lb per hour.

Steam for drying usually is conducted through tubes, over the external surfaces of which the material to be dried is passed. The tubes may be arranged immobile in towers or mobile in rotary driers. With 10 psi pressure in the tubes, the steam consumption is approximately 1½ lb per 1 lb of water evaporated.[12]

On the basis of these data, a simple, high-speed steam engine driving a centrifuge delivering 25 tons of material per hour would have enough exhaust steam to evaporate water equivalent to 3.3 pct of the weight of the centrifuged material. Referring to the previous discussion of Fig 2, such a quantity of heat would be small. It would be slightly greater than that required for the sensible heat needed for a 100°C rise in the temperature of the solid.

The problem, therefore, of exhaust steam from a steam prime mover, especially for one more efficient than the simple engine, is not that of a surplus but of a deficiency of drying steam. The various requirements however are adjustable, such as use of additional steam prime movers, or a steam-electric plant, for the total source of power.

Heat from a diesel prime mover can also be utilized,[4] however, the available exhaust heat is much smaller equivalent, after converting to steam, to evaporating only 0.1 pct, instead of 3.3 pct, of the weight of the centrifuged material as water. Again the requirements are adjustable such as by using additional diesel power for other drives, or by using the diesel-electric plant as the source of total power. If the exhaust gases were brought into direct contact with the material to be dried, more heat would be available. Objectionable carbonaceous matter and soot could be initially removed with electrostatic precipitators. The heat in the jacket cooling-water of the diesel may also be of value.

Incidentally, Overstrom[16] has described the use of steam in drying sand in preparation for his pyroelectric process which he actually carried into practice for years. His method was to cascade the sand over pipes through which steam was passed. Pipe is a common standard commodity and for this reason a cascade drier can be constructed with little expense.

LIQUIDS OTHER THAN WATER

Minerals can be separated according to their differences in dielectric constant when suspended in a nonconductive liquid dielectric of intermediate dielectric constant.[2] The method was originated by Hatfield[17] and described further by Holman and Shepherd.[18] Since the electric forces in this process are determined by volume properties instead of surface properties, as commonly encountered in conductivity and contact potential separations, surface impurities, unless of very high dielectric constant, have little influence on the separation. The effect of surface impurities in conductivity separations has been previously described.[19] Also because of these volume properties, the potentials, electric-field regions, and electrodes, can be changed dimensionally to match the particle size. Theoretically a wide range of particle sizes may therefore be separated.

Hatfield developed this process to a 1 ton per day stage working on Cornish cassiterite and Colorado roscoelite, both with siliceous gangues. The ores were separated at slime size, evidently the size needed for liberation. Apparently there is hardly any limit to the fineness of the slimes at which the process will operate.

With these slime sizes, Hatfield reports 30 pct of the dry weight as liquid retained in the filter cakes and recovers this by distillation at 300°C. Such an operation is much more expensive than removing water by evaporation in a direct-heat rotary drier. Any loss or incomplete recovery of the dielectric liquid—in this instance a solution of nitrobenzene in kerosene—will also result in prohibitive costs.

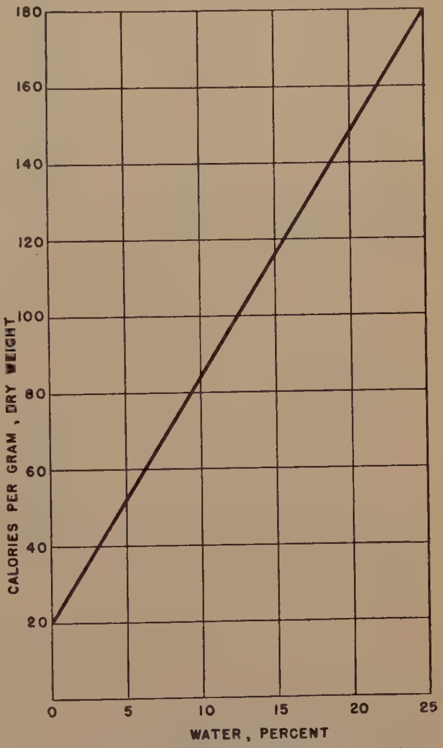

FIG 2—HEAT REQUIREMENT FOR DRYING. (FACTOR: MULTIPLY BY 3600 TO GET BTU PER TON).

No commercial applications of the process have been reported in the literature. It is believed that if the actual separation can be made to operate efficiently, the only obstacle to successful commercial application is solution of the dielectric liquid-recovery problem.

An alternative to the removal of the dielectric liquid by distillation, or centrifuging, is displacement as a liquid by water. There is no need for drying the rejected gangue. Drying a large amount of gangue merely represents a waste of energy. Displacement of oils by water is practiced in the recovery of the organic sink-float liquids for separating coal. However, the coal is in a much coarser size and also given a preliminary coating with a wetting agent. The problem here is to operate at much finer sizes and without the presence of liquid water on the particles before introduction into the liquid dielectric. Preliminary coating with a wetting agent usually necessitates the addition of liquid water to the mineral surface.

TABLE 4—*Water Displacement of Oil from Minus* 100-*Mesh Solids*

Test No.	Solid	Initial Treatment	Oil	Water Displacing Oil	Solid Added to Oil	Displacement of Oil by Water
1	calcite	none	kerosene	daxad[d]	dry	none
2	calcite	daxad[a]	kerosene	daxad[d]	moist	none
3	calcite	calgon[b]	kerosene	calgon[e]	moist	complete
4	calcite	calgon[b]	kerosene	calgon[e]	dried	complete
5	spodumene[f]	calgon[b]	kerosene	calgon[e]	dried	complete
6	spodumene[f]	none	kerosene	calgon[e]	dry	partial
7	spodumene[f]	none	kero-nitro[c]	calgon[e]	dry	complete

[a] Rolled with pebbles with 4-lb daxad (sodium lignin sulphonate), and 50-lb water, per ton.
[b] Same, but with 2.5-lb calgon (sodium hexametaphosphate), and 50-lb water, per ton.
[c] 4 parts kerosene plus 1 part nitrobenzene.
[d] Water contains daxad.
[e] Water contains 0.15 pct calgon.
[f] Also associated feldspar and quartz.

Dielectric liquids other than kerosene and nitrobenzene may be used. The requirements are low cost, low volatility and low toxicity. The hydrocarbon kerosene has a much lower surface tension than water, $\sigma = 72.8$. It compares with benzene, $\sigma = 28.9$, and hexane, $\sigma = 18.4$.[20] The function of the surface active agent is to lower the surface tension of the water so as to permit it to wet the solid and displace the oil. Kerosene has the lower dielectric constant, $\epsilon = 2$,[21] and is used in greater proportion for the lower dielectric-constant minerals. Nitrobenzene has the higher dielectric constant, $\epsilon = 36$, and is used in greater proportion for the higher dielectric-constant minerals. The surface tension of nitrobenzene, $\sigma = 43.9$, is higher than that of kerosene. Fortunately the lower dielectric-constant minerals are, in general, more easily wetted by water. Thus with air and water in equilibrium with quartz and galena, the angles of contact are < 31 and 86.[22] Quartz has the low dielectric constant of 5[23] and galena the high dielectric constant of > 80. An exception is calcite which has the low dielectric constant, $\epsilon = 8$,[23] and the high contact angle, $\theta = 70$.

The displacement of oils from calcite and spodumene, with associated minerals, by water with the aid of surface active agents is summarized in Table 4.

The general procedure consisted of adding the solid to the oil, removing the excess oil, adding this solid to the water, removing the excess water, and then determining the residual liquid by distillation. Excess oil and water were removed by absorption with filter paper. After the oiled solid was added to the water, the mixture was agitated violently and then allowed to stand. When oil was displaced the solid settled to the bottom and the oil rose and coalesced to a layer floating on the water. Mixtures up to and including 2 parts of nitrobenzene to 1 part of kerosene float readily on water.

The solids were added to the oil in three conditions, pretreated and moist, pretreated and dry, and not pretreated and dry. The surface active agent was also present in the displacing water. Sodium hexametaphosphate was effective in permitting the oil to be displaced by water even without pretreatment. The difference in tests 6 and 7 is explained by the fact that the nitrobenzene has a higher surface tension than kerosene. It is preferable to have the surface agents which lower the surface tension of water inorganic and water-soluble only, and the constituents in the oil, organic and oil-soluble only.

The rock probably could be pretreated in a dry state by rolling or agitating with the solid or concentrated surface-active agent. That the agent will spread over the rock in this condition is demonstrated in the use of grinding aids for dry-grinding Portland cement clinker.

In test 6 the partial displacement of oil by water may indicate preferential wetting of one mineral constituent with water. Liquid water has a high dielectric constant. Such an increase by water filming in the over-all dielectric constant could aid the dielectric separation of the minerals. It could also aid in conductivity and contact-potential separation.

If the dielectric liquid has some solubility in water, the displacement water can be recycled and recovered from the rock by washing with fresh water. This procedure also reduces the consumption of the surface-active agent. Drying of the dielectric liquid after recovery may be necessary.

In arid regions, displacement of the dielectric liquids with water is not practical. However, there are other distillation alternatives. One alternative is the use of low-boiling point dielectric liquids to increase the variety of available heat sources. Another alternative is distillation under conditions in which the heat changes are maintained in a closed cycle. Here the heat required for distillation is supplied by the heat evolved in the dielectric liquid or water recovery by condensation.

Since a temperature drop results from heat-transfer requirements, some additional energy must be introduced into the system. This additional energy W is only a small fraction of the total heat Q_2 involved in the distillation, provided the temperature drop for heat transfer is also small. It is represented by the familiar thermodynamic relationship for the heat engine,

$$\frac{W}{Q_2} = \frac{T_1 - T_2}{T_2}$$

where T_1 and T_2 are the absolute temperatures respectively for the condensation and the distillation. The energy W would be supplied by a compressor, probably of the rotary type, operating in the pressure range of the boiling points T_1 and T_2.

A convenient type of distillation apparatus could consist of a jacketed and closed trough with a semicylindrical bottom, along which the material is conveyed by a spiral agitator. The jacket serves as a condenser for recovering the liquid. The compressor operates on the pressure differential between the inside of the trough and the jacket. Waste heat also has its applications here when the thermodynamics of the problem are considered. It would be useful in supplying the sensible heat for the feed.

SUMMARY

In the electrostatic separation of broken solids by use of (1) differences in conductance, (2) differences in contact potential or frictional electricity or (3) differences in dielectric constant, the effects of the presence of moisture or other liquid in various proportions is discussed. The range of permissible moisture content varies with the minerals involved and the process used. Certain of them are shown to demand very low moisture content on the mineral particles and in the ambient air; in other instances, high concentrations of moisture have little effect.

Water present in volumes greater than permissible can be removed mechanically as liquid or by evaporation. The centrifuge is capable of removing water mechanically to very low values on coarsely sized material. This is shown by the electrostatic separation of fluorite from siliceous gangue, and of feldspar from quartz, after the mechanical removal of liquid water with the centrifuge.

Cost estimates for removal of water with the centrifuge and by evaporation indicate that the centrifuge is less expensive. Use of steam and diesel prime movers for driving the centrifuge, with the application of the exhaust steam and waste heat for final drying, has possibilities.

Rock mined at a coarse size contains very little moisture, even if drenched. Drying for the electrostatic separation of coarsely mined rock consequently does not require the evaporation of much water.

The recovery of dielectric liquid medium in the dielectric-constant method of separation presents the same problems but can be solved in a different manner. Here, displacement with water is more satisfactory. This method can be made to operate on material of very small particle size.

ACKNOWLEDGMENT

Acknowledgment is given to O. C. Ralston, whose support has made this research possible, to J. B. Zadra and V. H. Gottschalk for administrative guidance and encouragement, to J. E. Conley for centrifuge cost estimates, to H. W. Jaffe for petrographic examination of the fluorite-siliceous gangue, and to F. D. Lamb for furnishing the mineral samples.

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DISCUSSION

H. L. BULLOCK*—This paper is a valuable addition to the literature pertaining to the art of electrostatic separation. It is a triple contribution to the art, for the following reasons:

First, the reports of the tests which he made prove that certain mineral separations are possible at high humidity and with materials containing an appreciable amount of water.

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Successful separation of these materials under almost run-of-the-mine conditions indicates that, as new problems of separation arise, careful tests must be made before electrostatic separation is rejected as a possible method, due to the old belief that successful operation was only possible with dry materials under low humidity conditions.

Second, the information included in his paper on the costs of mechanical and thermal drying, on the amounts and types of reagents necessary for surface treatments, and on the methods of recovery of dielectric liquid mediums will allow much more accurate estimates of costs of electrostatic separation of minerals and other substances.

Third, once again, as in his joint paper with O. C. Ralston in 1940,[24] Mr. Fraas has carefully set forth the relative nature of the surface charges and the effects of variation of surface and mass conductivity. Such scientific approach should relieve us of any faith in the Johnson Mineral Table of the electrostatic behavior of various minerals which may be still existing and teach us the proper approach to electrostatic separation problems. On a practicality basis, Mr. Fraas has restricted himself to separations which occur in actual mining practice. This is a distinct improvement on earlier publications by others which stress such noncommercial separations as graphite from sulphur[3] and graphite from corundum.[25]

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The Electrostatic Separation of Several Industrial Minerals

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INTRODUCTION

ELECTROSTATIC methods of separation are used only when some peculiar advantage is gained. Such cases are minerals that are not separable by differences in specific gravity or magnetic response and that are desired in particles as large as 10-mesh, a size on which flotation separation is inefficient. This paper covers several separations of this type and others. It includes studies in the electrostatic behavior of kyanite, sillimanite, and zircon, simple silicates that cannot be called either acid or basic but are more nearly neutral. Kyanite and sillimanite are both desired in 10-mesh size and are difficult to beneficiate from low-grade ores to give that type of concentrate. A separation of garnet from diopside was studied because this mixture had been removed from a wollastonite product by drying and magnetic separation, and no special or extra-preparation for electrostatic separation was necessary.

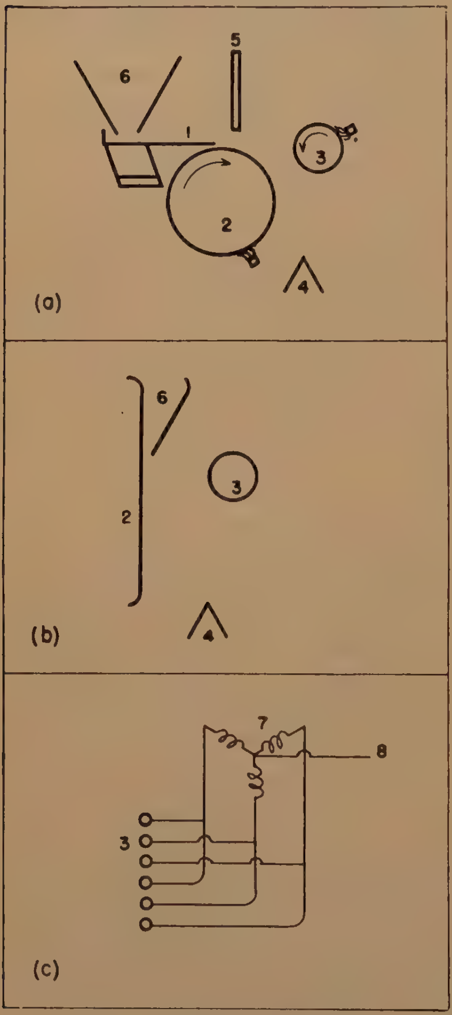

FIG 1—TYPES OF SEPARATORS.

THE SEPARATOR

The separator used in these tests is of the nonhomogeneous field type shown in Fig 1*a*. It consists of a 11.4-cm-diam rotating metal roll *2* at ground potential over which the dry mixture of minerals is fed. The roll serves as one electrode of the separating field, and the other electrode is that of *3*, a smaller 5-cm-diam rotating metal cylinder which is held at a high direct-current potential. In this paper the term "electrode" will be confined to the electrode at *3*, roll *2* being designated only as "roll."

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This arrangement of field is widely used, exemplified by the Johnson separator. Material on roll *2* can take on a charge of the roll polarity by conduction or by other means of imparting charge. Frictional electrical charging can be added to this separator by the vibrating aluminum plate *1*, which serves also as a feeder. These vibrating "electrifiers" have been described in a previous publication.[1] Charges caused by frictional electrification may superpose on or oppose charges taken on by conduction. If frictional charging is negligible, the vibrating plate functions only as a feeder or may be eliminated. The dividing edge *4* provides for the split into two fractions. The fraction adjacent to the roll side is designated the nondeflectate. A feed hopper is shown at *6*, and *5* is a plate for electric shielding. In simulating a multiple-stage or multiple-roll separator, one of the separated fractions was merely repassed through this single stage, the number of passes corresponding to the number of rolls. Both the separator and material were heated.

Other types of separators can also be used. One alternative is the immobile cascade type of separator illustrated in Fig 1*b*. Here roll *2* has been replaced by a vertical plate. A second parallel plate could be substituted for electrode *3*, but it tends to obstruct the deflected particles. There are no moving parts. Although this type of separator has a lower separation efficiency for some minerals than that of Fig 1*a*, it is simpler and less expensive to construct. Sheet *6* can serve not only as a feed hopper but also as the more important function of a baffle for directing the stream of feed.

With minerals which tend to adhere, it is necessary to tap or vibrate plate *2* and electrode *3*. A self-cleaning grid electrode, consisting of a planar arrangement of round tubes, is illustrated in Fig 1*c*. The tubes are connected in successive order to a polyphase alternating-current, high-voltage source *7* having a downward sweep in phase variation. The alternating-current source system is also connected to the usual higher-voltage direct-current source *8*. This electrode may serve as a replacement for the electrode *3* in Fig 1*b*.

Other alternatives are the vibrating-table-type separator described in a previous publication,[2] and a simplification of the separator in Fig 1*a* by a replacement of roll *2* with the discharge edge of vibrating plate *1*.

PRELIMINARY CONDITIONING

Some minerals can be separated without preliminary conditioning; others require it. Conditioning procedures may be divided into two groups; those which result in (1) surface cleansing, or in the removal of certain constituents from the mineral surfaces, and (2) those which form surface films. The first group usually requires an aqueous pulp, unless the constituents removed can be volatilized, abraded, or dispersed. The second group can proceed in either a wet or dry state.

The mechanism of surface cleansing is more complicated than is generally assumed. Some light has been thrown on the subject by research in fields other than electrostatic separation. For example, surfaces which are exposed to a low-pressure electrical gas discharge, the ionized part of a gas flame, sparks in air, or to a high-frequency brush discharge, are changed so that moisture later deposited on the surface from a humid atmosphere forms a continuous transparent film. In contrast to this, the same moisture deposited on the untreated surface forms the customary minute-droplet fogging type of deposit. If the object is a plate of glass and if the cleaned portion is in the shape of a design or figure, the design can be transferred to another plate of glass merely by bringing the second plate in proximity to the first plate, but not actually in contact. With glass, the cleaned portions have a greater electrical conductivity and a greater coefficient of friction.[3] The electrical gas-discharge treatment also increases the adhesion of metal films formed by high-vacuum deposition.[4] This increased adhesion suggests other applications to the processing of industrial minerals, such as the metal filming of abrasive powders so that they may be compacted later by powder metallurgy methods into useful cutting tools. The same technique could be extended to the metallic and nonmetallic filming of other powders to produce desirable surface properties. These properties may contribute color and reflectivity, form electrical insulating films such as on powders for magnetic cores, or, for metal films on metal powders, provide for certain powder metallurgy objectives.

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[1] References are at the end of the paper.

For aqueous conditioning, the minerals in the acidic igneous rock frequently respond to any of the highly-dissociated acids, such as sulphuric or hydrochloric. The basic pegmatites usually require hydrofluoric acid or a mixture of sulphuric acid and a fluoride salt. Minerals that are heavily iron-stained can be cleaned with the more effective and less expensive sulphurous acid or its highly reducing derivative, zinc hydrosulphite. Attrition scrubbing is satisfactory for the removal of friable films.

Benzoic acid has been consistently used by us for selective film formation. Although previous applications of benzoic acid were confined to the more basic minerals, such as ulexite[1] and zoisite,[5] the work presented here is an extension to the less chemically reactive or neutral types, such as kyanite, sillimanite and zircon. Also in previous work an effort was made to apply a filming reagent either as a gas or in a dilute solution (followed by drying). The present investigation demonstrates that the solid or liquid reagent, undiluted, may be added directly to the minerals to be conditioned.

One advantage in the use of benzoic acid for film formation is that it is volatile. The use of a vapor avoids the possible error of adding an excess of reagent. Additional tests have shown that other organic acids having a hydrocarbon group of sufficient length are also satisfactory. These include salicylic, phthalic, heptylic, and cresylic acids. Benzoic and salicylic acids contain carboxyl groups attached to aromatic rings. Heptylic acid contains a carboxyl group with an aliphatic chain. Cresylic acid is a mixture of cresols and xylenols which contain hydroxyl instead of the acid-carboxyl groups. However, the hydroxyl group when attached to an aromatic has acid properties. Oxalic acid was not satisfactory. This can be explained by the fact that oxalic acid does not have a hydrocarbon appendage of sufficient length.

In the aqueous conditioning of small batches, the slurry was mixed with the aid of a bakelite rod in a beaker (bakelite for HF) slowly rotating about its cylindrical axis, which was fixed at an angle with the vertical. Larger batches were agitated in a rectangular box by a slow, rocking motion. Washing was performed by the customary decantation method and was followed by drying.

Conditioning with benzoic acid was done in a pyrex glass beaker with rod. The anhydrous acid was weighed on a sensitive balance and introduced into the hot, dry solid by stirring and rolling.

GEORGIA SILLIMANITE

The separation of a Georgia sillimanite schist containing sillimanite, quartz and biotite shows where conditioning in both the aqueous and dry state is beneficial. Aqueous-acid treatment results in a change from practically no separation to a good separation, while additional treatment of the dried product with benzoic acid yields further improvement. For example, two tests, identical except for the omission of benzoic acid in one, yielded respectively for grade and recovery of sillimanite, 93 and 82 pct without benzoic acid, and 96 and 89 pct with benzoic acid.

Better separation is obtained with the use of hydrofluoric acid in the aqueous-acid conditioning than with other inorganic acids. However, the addition of only minute amounts of hydrofluoric acid to the other inorganic acids is satisfactory. A separation utilizing this combination is summarized in Table 1. The separator was that illustrated in Fig 1*a* with the attached vibrating plate electrifier. The electrifier may introduce frictional charging. A minus 14-plus 65-mesh rougher wet-gravity table product was conditioned for 20 min. with 2.5 lb of H_2SO_4 and 0.25 lb of HF per ton of feed in a 65 pct solids pulp. After washing and drying by heating, it was further treated by stirring 1 lb of benzoic acid per ton into the hot, dry feed, the acid vaporizing and spreading through the mass. Conditioned in this manner, the quartz is deflected and the sillimanite not deflected when the electrode is positive, and vice versa when the electrode is negative. The biotite is deflected with either polarity. On passage over a 3-roll separator with the adjacent electrodes of negative polarity, the nondeflectate was passed over succeeding rolls to finally yield the rougher tailing. The deflectates were combined as heading for a second 3-roll separator having positive electrodes. The nondeflectate passed over succeeding rolls to yield, after the last roll, the sillimanite concentrate. The combined deflectate from this second separator constituted the reclean tailing.

For market requirements a sillimanite concentrate having less than 5 pct quartz is satisfactory. The concentrate of Table 1 meets this requirement. The main residual impurity, biotite, can be removed by passage over additional rolls or by a magnetic separator.

TABLE 1—*Separation of Georgia Sillimanite*

Product	Weight Per Cent	Composition, Per Cent[a]			Per Cent of Total Sillimanite
		Quartz	Biotite	Sillimanite	
Rougher tailing	41.7	91.5	5.0	3.5	4.0
Reclean tailing	27.2	55.6	26.0	18.4	13.8
Recleaned concentrate	31.1	1.4	2.9	95.7	82.2
Composite	100.0	53.7	10.0	36.3	100.0

[a] By acetylene tetrabromide sink-float and alnico magnet for biotite.

BAKER MOUNTAIN, VIRGINIA, KYANITE

Kyanite-bearing rock with quartz-muscovite-chlorite gangue responds in a similar manner to aqueous and dry conditioning. Without conditioning the separation is negligible. In the aqueous conditioning inexpensive acids such as sulphuric are satisfactory and there is no appreciable increased separation efficiency through the use of hydrofluoric acid. A typical separation is summarized in Table 2.

A minus 10-plus 65-mesh fraction of partly weathered Baker Mountain, Va., rock, from which clay had been previously removed by washing, was conditioned for 20 min. with 10 lb of H_2SO_4 per ton of feed in a pulp containing 65 pct solids, washed and dried by heating, and further treated, while hot, with 2 lb of benzoic acid per ton. The electrostatic flowsheet, involving two 3-roll separators of the Fig 1*a* type, was the same as that for the sillimanite. The quartz, chlorite, and muscovite are deflected and the kyanite not deflected when the electrode is positive, and vice versa when the electrode is negative. The rutile and ilmenite are deflected with either polarity. The nondeflectate from the first separator with negative electrodes constituted the rougher tailing, the nondeflectate from the second separator with positive electrodes, the kyanite concentrate, and the deflectate from the second separator, the reclean tailing.

TABLE 2—*Separation of Baker Mountain, Va., Kyanite*

Product	Weight Per Cent	Sink-Float and Petrographic Assay, Per Cent				Per Cent of Total Kyanite
		Float	Sink,[a]			
			Kyanite	Hematite	Others[b]	
Rougher Tailing	58.1	94.5	1		4	3
Reclean Tailing	18.3	51.4	10		39	8
Recleaned Concentrate	23.6	3.3	86	11[c]	0	89
Composite	100.0	65.1	23		12	100

[a] Acetylene tetrabromide sink-float assay with petrographic assay of the sink fraction.
[b] Rutile, ilmenite and hematite.
[c] Hematite stain on the kyanite.

Unfortunately this sample is heavily iron-stained, so that the impurity in the concentrate assaying 96.7 pct sink is a hematite coating. However, the data show what can be accomplished with the mechanically free constituents. The hematite coating does not seem to affect the separation, since both the blue-unstained and the brown-stained kyanite particles collect in the same fraction. Rutile and ilmenite are concentrated with the gangue in the reclean tailing. The minerals removed as gangue are chlorite, muscovite and quartz, with quartz predominant.

The use of benzoic acid increases both the kyanite recovery and the percentage of kyanite in the concentrate. The increase in recovery is clearly illustrated by the assay of the coarse sizes of rougher tailings of two tests summarized in Table 3. The two tests are identical, except in the use and omission of benzoic acid. For example, in the minus 14-plus 20-mesh tailing fraction, the kyanite content is reduced from 9.3 to 0.5 pct by the benzoic-acid treatment.

TABLE 3—*Assay of Sized Fractions of the Baker Mountain Kyanite Rougher Tailing*

Sized Fraction	Composition, Per Cent Kyanite	
	Test without Benzoic Acid Treatment	Test with Benzoic Acid Treatment
Minus 10 plus 14 mesh (a)*	27	3.7
Minus 14 plus 20 mesh (b)	9.3	0.5

* The weight per cent of the total tailing in fractions (a) and (b) were respectively 4.1 and 13.2 in the test without benzoic acid, and 3.7 and 10.2 in the test with benzoic acid.

NEW YORK GARNET—DIOPSIDE

This is an example of a mineral combination that can be separated without preliminary conditioning with reagents. The garnet and diopside are associated with wollastonite, the occurrence and uses of which have been described in a paper by Broughton and Burnham.[6] They give data for the separation of the garnet and diopside, as magnetic minerals, from the wollastonite, as nonmagnetic, adding that the garnet and diopside are magnetic enough to be attracted with an Alnico permanent magnet. The magnetic method will not yield a separation of garnet from diopside. A sample was accordingly submitted to the Bureau of Mines by K. D. Burnham, with inquiry as to an electrical method of separation.

With the conductivity electrostatic method, the garnet has a high probability of deflection. The probability for the diopside is low. Investigation shows that the vibrating plate electrifier has no influence on the separation (no frictional electric charging) and that a simple roll-type separator suffices. The quantity of garnet deflected from the grounded cylinder depends on the polarity of the charged electrode. At a potential at which 80 pct of the garnet is deflected after 3 passes with a negative electrode, only 45 pct is deflected with a positive electrode. This indicates that the garnet is a semiconductor with asymmetrical conductivity.

The importance of semiconductor properties in electrostatic separation has been discussed in a previous paper.[7] By constructing the separator roll of semiconductor material instead of metal, further advantage may be taken of these semiconductor properties. The asymmetric conductivity, most of which is controlled by the particle-roll interface, will depend not only on contact potential differences but also on whether the particle and roll are electron or positive hole conductors. For example, zinc oxide is an electron conductor and cuprous oxide a positive hole conductor. In positive hole conductors the current is carried by the transport of electron-missing places in the crystal structure instead of by electrons. Data on the influence of the type of semiconductor and the contact potential on the asymmetrical conductivity at the interface may be found in the cited literature of a previous publication.[7]

The concept of an asymmetrical transport of charge may possibly be extended to substances, intermediate between semiconductors and ideal nonconductors, where the transport may occur only at the interface without any additional transport deep within the internal structure.[8] For example, if a particle is in contact with a roll having an adjacent negatively charged electrode, the roll is positive and the particle negative at the roll-particle interface. The negative charge on the particle at this point is an induction charge, which is accompanied by an equal and opposite charge at the opposite extremity of the particle. If a transport of charge can occur across the interface, the charge on the opposite extremity alone remains, and the particle is deflected. An asymmetrical transport results when, under the conditions of a reversed polarity of the adjacent electrode, there is no transport and consequently no deflection. The polarized deflection of such apparently nonconductive particles has previously been explained on the basis of an initial frictional electrification. Although frictional electrification has been demonstrated[1] for some separations, no experimental data are yet available to substantiate the alternative mechanism.

The high probability of deflection of the New York garnet is in contrast to the low probability of the garnet in Oregon beach sand. A difference in its iron content is evidently an influential factor. The Oregon garnet is the low-iron, high-calcium grossularite, whereas the one now discussed is the high-iron andradite. An analogous increased probability of deflection with increased iron content is exhibited by high-iron Oregon chromite as contrasted with the low-iron, high-magnesia African chromite.[7]

When New York garnet is separated electrostatically from diopside without pretreatment, the garnet is recoverable in a concentrate assaying 1 pct diopside, with approximately 95 pct recovery. Small amounts of wollastonite remain with the diopside. If this wollastonite is discarded magnetically, the electrostatic-diopside concentrate would assay approximately 90 pct diopside. The diopside content of the heading is only 5 to 10 pct. Approximately 64 pct of the diopside is recovered in the concentrate, and 20 pct remains in the reclean tailing. It is probable that the reclean tailing can be reworked for further recovery of diopside. Diopside is normally of no value but promising ceramic uses have recently been described.[9]

These results are illustrated in the electrostatic separation of a minus 10-plus 20-mesh magnetic concentrate summarized in Table 4. A closely sized feed resulted from a preliminary magnetic separation of garnet-diopside from wollastonite. Such a close sizing is not necessary for the electrostatic separation. Both the material and separator were heated. The feed was passed through a 3-roll separator, with the nondeflectate passing to succeeding rolls. The deflectates from the 3 rolls were combined and repassed through a second 3-roll separator. The deflectates from the second 3-roll separator constituted the clean garnet concentrate. The final nondeflectate from both 3-roll separators were combined and passed over a 9-roll separator, the deflectates from this constituting the reclean tailing and the final nondeflectate the diopside concentrate.

TABLE 4—*Separation of New York Garnet-Diopside*[a]

Product	Weight Per Cent	Composition, Per Cent				Per Cent of Total	
		Garnet	Diopside	Wollastonite	Locked Grains	Garnet	Diopside
Garnet concentrate	90.0	74	1[b]	0	25[c]	97	17
Diopside concentrate	4.3	5	80	11	4[d]	0	64
Reclean tailing	5.7	33	19	4	44[e]	3	19
Composite	100 0	69	5			100	100

[a] Minus 10-plus 20-mesh heading.
[b] One-half consists of garnet-diopside locked grains.
[c] Garnet-wollastonite locked grains.
[d] 60 pct garnet-wollastonite and 40 pct diopside-wollastonite locked grains.
[e] 80 pct garnet-wollastonite, 15 pct garnet-diopside-wollastonite, 5 pct diopside-wollastonite locked grains

All the stages utilized negatively-charged electrodes and also carried the same rate of feed. Since the 3-roll separators remove 88 pct of the feed weight, the 9-roll separator can be of much smaller units. Only 12 pct of the capacity of the 3-roll separators is required.

At the 10-mesh size a large proportion of the wollastonite is free of the magnetic garnet and diopside. Broughton and Burnham state that grinding to minus 35-mesh is necessary for complete liberation. The recommended flowsheet is to grind to a coarse size, separate a portion of the feed weight magnetically as free wollastonite, grind the residue to minus 35-mesh, remove dust, separate the remaining wollastonite magnetically, heat the magnetic "residue," and separate into garnet and diopside by electrostatic means.

The advantage of the electrostatic method in this case is that there are no reagent costs and no drying costs. The heating for the electrostatic separation is merely to remove adsorbed moisture, which is negligible compared to the evaporation of liquid water in drying. Water that may be accumulated in mining is small because of the small surface:mass ratio of the coarse rock. The electrostatic separator, once adjusted, requires very little attention. There is no constant determination and adjustment of reagent flow and pulp densities.

FLORIDA KYANITE—ZIRCON

The west Florida beach sands have rounded grains, which are confined within a narrow range of size limits. Because of the same rounded-grain shape of all the mineral constituents, the quartz can be separated from the kyanite and zircon by such gravity methods as wet tabling. Crushed-rock kyanite, as represented by the previously described Baker Mountain, Va., sample, has an elongated grain. This elongated shape nullifies the advantages of density differences during tabling and necessitates application of other methods, such as the electrostatic. The same electrostatic procedure that was satisfactory for the Baker Mountain kyanite is also applicable to the Florida beach sands, with the added fact that the separation is more efficient. The increased efficiency eliminates the necessity for using benzoic acid.

The electrification characteristics are intense enough apparently to pass beyond the intensity for good separation. After conditioning with dilute sulphuric acid, washing and drying, excellent separation is obtained at atmospheric temperature in air of moderate humidity. However, if the dry sample is heated—a procedure that usually improves most separations—agglomeration ensues upon stirring or agitation. This agglomeration is detrimental to electrostatic separation of the quartz.

The appearance of the agglomeration is somewhat similar to that resulting when the sand is wetted. If the quartz is separated from the kyanite and zircon, agglomeration does not occur in either of these two separated fractions. Since agglomeration requires intimate contact of the quartz and the remaining constituents, it is concluded that frictional electrification is the charge-producing mechanism. Polarities observed during separation indicate that the quartz has a negative charge, the remaining constituents a positive charge.

This phenomenon of agglomeration is not unusual, since it is found to occur in some other mineral mixtures. Other mechanisms of charging are also possible. Charging may result from the formation of dipoles, such as by the pyroelectric effect.[2] Loeb, working for Overstrom, found that when a special variety of quartz sand was heated in a glass beaker with a glass stirring rod, a sudden agglomeration occurred above a certain temperature. The grains of sand adhered together in chains, similar to a series of magnets, indicating the formation of dipoles. At the Titanium Alloy Manufacturing Co., clean zircon which has been electrostatically separated from rutile remains agglomerated and difficult to pour for nearly 24 hr.

The agglomeration of the quartz-kyanite-zircon disappears on exposure to an alternating-current corona. For example, when the sand was passed over a plate above which, at a 7-cm distance, were placed needle points excited with a 15-kv (rms), 60-cycle alternating potential, the agglomeration was rapidly dissipated. Gas flames and other ionizing atmospheres are also satisfactory. Such a charge-neutralizing method would immediately precede the separation step. It would probably also improve other separations by initially removing undesirable electrical surface charges.

Agglomeration of the quartz-kyanite-zircon results only after conditioning with highly-dissociated acids, such as hydrochloric and oxalic. It does not occur after treatment with boric acid, sodium hydroxide, sodium hexametaphosphate or attrition cleaning. After storage in an open vessel for 12 days, the agglomeration fails to reappear on heating and stirring, but it does reappear after storage in a closed container for one year.

If the agglomerated material, while still hot, is treated with approximately 2 lb of lauryl amine per ton, the agglomerating phenomenon disappears, and the material can be stirred with remarkable ease. Measurements have shown that the torque required for stirring after amine treatment is only 61 pct of that required for the original beach sand. The greater torque with the original sand is probably caused by the alternate positive and negative charges at the crystal surface, a result of atomic arrangement in an ionic crystal. Filming results in an outer layer which is completely of one polarity at the outer surface extremity. Amine filming also increases the apparent electrical surface conductivity.[1] An unrestrained movement of electrical charge may be influential in reducing dipole attraction to yield a low-friction surface similar to that obtained with vacuum-deposited metals, such as barium.

TABLE 5—*Separation of Florida Kyanite-Zircon*

	Weight Per Cent	Assay, Per Cent[a]				Per Cent of Total	
		K	Z	Q	O	K	Z
Kyanite concentrate	63.6	96	2	1	1	84	6
Zircon concentrate	18.8	12	81	1	6	3	66
Reclean tailing	17.6	55	36	2	7	13	28
Composite	100.0	73	23	1	3	100	100

[a] K, Z, Q, O = kyanite, zircon, quartz, others (ilmenite, rutile, staurolite, and quartz).

After amine treatment there is no decrease in the separation efficiency of the quartz; however, later separation of the kyanite-zircon fraction is less efficient.

The separation of kyanite from zircon is improved by conditioning with benzoic acid in a manner similar to that described for sillimanite and crushed-rock kyanite. In the example summarized in Table 5, the sand was conditioned for 20 min. with 0.5 lb of HF and 50 lb of H_2SO_4 per ton of feed in a 60 pct solids pulp, followed by washing and drying. After removal of the quartz by the electrostatic method in a manner similar to that for the Baker Mountain sample, but without benzoic-acid treatment, the hot, dry solid was conditioned with 1 lb of benzoic acid per ton to activate zircon. Electrostatic separation consisted of first passing the sand through a 2-roll separator, each roll having vibrating plate electrifiers and negative-active electrodes. The nondeflectate passed over succeeding rolls to yield the kyanite concentrate as the final nondeflectate. The zircon-bearing deflectates were combined as heads for a second 3-roll separator having positive electrodes. In this second separator (with reversed potential), the combined deflectate constituted the "reclean tailing" (or middling), and the final nondeflectate the zircon concentrate.

This process yields an 84 pct recovery of kyanite, with a 96 pct grade. The zircon concentrate is not satisfactory for marketing, since it has only an 81 pct grade. However, it is possible that both the grade and recovery could be increased by further passage of the concentrate and reclean tailing. There are many factors still to be investigated. One of these is the gradual disappearance of the effect of the benzoic-acid conditioning upon continued passage through the separator. The film either evaporates or wears off. This suggests enclosing the separator in a compartment carrying a partial pressure of benzoic-acid vapor or the use of a less volatile reagent. More thorough investigation of the frictional electrical charging and the selection of a plate electrifier with the proper contact potential may also contribute to further improvement. These phenomena have been baffling but are detailed above to show some of the unsolved problems of the various electrostatic-separation processes and to advance discussion of the causes of variations.

COMMERCIAL APPLICATION

The Georgia sillimanite and the Florida kyanite-zircon are obtained as rougher concentrates by wet tabling, and the

Virginia kyanite by washing for clay removal. These wet concentrates are ideally adapted to aqueous conditioning preliminary to drying before electrostatic separation.

The quantities of reagents specified are arbitrary and not the minimum. Further tests of the continuous pilot-plant type would yield definite values. Unusually large quantities of acid, such as the 50 lb per ton in Table 5, would not be consumed and would necessarily be recirculated. Observation of the evaporation and sublimation of the benzoic acid during the open-vessel conditioning indicates that this reagent was added in excess for the quantities given in the preceding text. For continuous operation, a closed vessel of the rotary-drum type with small feed and discharge openings would conserve the volatile reagents.

No cost estimates have been made, since it is believed that there are possibilities for further improvement in some of the present practices. For example, in the drying by means of heat of a product dewatered in a rake classifier, the cost will be high if a large amount of gangue is present. However, by centrifugal dewatering preliminary to final drying with heat, there should be an appreciable reduction in drying costs. This is described in another paper.[10]

When the initial cost of the electrostatic separator is an appreciable part of the beneficiation cost, it is advantageous to make a rough concentrate in a simpler and less expensive type of separator, such as the immobile cascade type illustrated in Fig 1*b*. The rougher concentrate from this step then would be recleaned in the more efficient and more expensive type of separator.

SUMMARY

The electrostatic separation of sillimanite, kyanite, andradite, diopside, and zircon from commercial samples of rocks has been described, as illustrative of what may be accomplished with other combinations of minerals. The information thus presented includes particle-surface conditioning and the adaptation of each separation to a particular type of separator. Associated electrical phenomena may have future applications to industrial minerals. These include the compacting of abrasive powders and the agglomeration and dispersion of other powders. The electrostatic method has its advantages in size ranges close to 10 mesh, where other methods fail, and frequently in the elimination of reagents and excessive moisture.

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J. B. Clemmer, formerly of the Tuscaloosa Station, submitted the sillimanite wet-table concentrate which was prepared under his direction. Acknowledgment is also given to the Zirk Corporation for supplying samples of Florida beach sands.

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DISCUSSION

Shiou-Chuan Sun*—It is indeed a pleasure to read an additional accessible publication on electrostatic separation written by Drs. Fraas and Ralston who have achieved so much in the research work of this field during the past several years. This paper illustrates that, under suitable conditions, the process of electrostatic separation of minerals can be advantageously utilized. It also serves to advance the technique of preliminary conditioning of minerals with chemical reagents for better separation.

The actual impressed voltages and the horizontal and vertical locations of the dividing edge of the separator, used in the separation of the different ores shown in Tables 1 to 5, were not reported in this paper. In my opinion, these points should be clarified for the convenience of other research workers and practical applications.

The experimental data were achieved by means of a Johnson type separator in conjunction with an aluminum vibrating electrifier. "Charges caused by frictional electrification may superpose on or oppose charges taken on by conduction." Without some parallel experimental data of the same ores tested in the absence of the aluminum vibrating plate, it is not only difficult to ascertain whether the aluminum vibrating plate functioned as an electrifier or merely as a feeder, but also impossible to determine the relative magnitude and the mutual effect between the frictional and the conductive charges in the separation of different ores used in this paper. Consequently, we can not decide whether the aluminum vibrating plate or even the frictional charge is a necessary factor for the separation of any or all ores.

This paper is a valuable addition to the field of electrostatic separation. The authors deserve to be heartily congratulated.

F. Fraas and O. C. Ralston (authors' reply)—With respect to Dr. Sun's inquiry, a detailed study of apparatus and source of electrical charge is beyond the scope of this paper which was merely for the purpose of encouraging the commercial application of the electrostatic method. However, we are planning a later paper relating the source of charge to the contact potential, dielectric constant, conductivity, and ionic or electron emissivity.

The low conductivity and low dielectric constant minerals, quartz, kyanite, zircon, and sillimanite, separate efficiently in a few stages of the simple roll type separator. In contrast, minerals which have a high conductivity or high dielectric constant, and which separate on the basis of these properties, usually require a large number of stages in the simple roll type separator. Examples are ilmenite, rutile, and chromite.

More efficient separation of the high conductivity minerals was obtained with a corona type separator consisting of a rotating copper roll and a rotating dielectric electrode. Corona charging yields an overall average of the surface conductivity of the particle. The variations resulting from minute surface irregularities and possibly the high dielectric constant effect of our reference 8, are eliminated.

The dielectric electrode principle has been described by Bullock to whom reference is given in a previous paper. However, the corona needles or wire which Bullock uses for the purpose of charging only the dielectric electrode is adapted by us to also charge the mineral particles. Adjustment of the corona with respect to the relative charging of the electrode and particles yields excellent separations of ilmenite, rutile, and leucoxene from staurolite, zircon, and quartz. The usual sulphuric and hydrofluoric acid conditioning improves the separation.

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In our estimation reagent conditioning is an important factor in securing the efficiencies required for commercial operation. The effectiveness of only traces of hydrofluoric acid is apparently related to its affinity for silica. Titanium[11] may have a similar response as a dry conditioning reagent.

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[1] W. A. Wooster and G. L. Macdonald: *Nature* (1947) **160**, 260.

Comparison of Galena and Ferrosilicon in Heavy-media Separation

By E. H. Crabtree, Jr.,* Member AIME

(New York Meeting, March 1947)

The heavy-media separation plant at the Central mill of the Eagle-Picher Mining and Smelting Co., near Picher, Okla., was started in February 1939. Since that time twenty-four million tons of lead-zinc ore has been treated in the mill, of which approximately 70 pct has been treated by heavy-media separation.

Until January 1945, galena concentrate from the flotation plant was used as the sink-float medium. At that time minus 100-mesh ferrosilicon was substituted for the galena, and its use has been continued. The purpose of this paper is to compare the operating results and the costs of operation of the two media. For the purpose of this comparison the period Jan. 1, 1943, to Jan. 29, 1945, has been taken to represent the operation using galena medium; the period Jan. 29, 1945, to Sept. 1, 1946, to represent the ferrosilicon medium. During the first of these periods about seven and one-half million tons was milled; during the latter, or ferrosilicon period, more than five million tons was milled. It is believed that these tonnages are large enough to give representative data.

Flowsheet

The Central mill has a capacity of 15,000 tons per day. Briefly, the treatment used consists of crushing the ore through 1½-in. and wet-screening out the minus 3/16-in. portion. The minus 1½ plus 3/16-in. washed product is treated by the heavy medium; the minus 3/16-in. product is deslimed and jigged. The heavy-media cone tailing is discarded and the cone concentrate is recrushed and jigged for the production of a coarse lead concentrate. The cone concentrate, impoverished of lead, is then ground in ball mills and treated by differential flotation. The minus 3/16-in. jig product is similarly treated.

Heavy-media Plant

The heavy-media cone plant treating the minus 1½ plus 3/16-in. product consists of one open-top cone 20 ft in diameter by 20 ft deep, equipped with a central airlift. Concentrate is discharged from the top of the airlift to a 3 by 14-ft low-head Allis-Chalmers drainage screen. Medium drains back into the cone, from the first half of this screen. The last half of the screen is provided with washing sprays for further removal of medium. Cone tailing is similarly drained and washed. Undersize from the washing screen is thickened and then cleaned in the medium-cleaning plant.

When galena was used as a medium, cleaning was done in Fagergren and Denver flotation machines. The ferrosilicon is cleaned in three 36-in. and one 24-in. Crockett magnetic separators. A flowsheet of the cone plant is shown in Fig 1.

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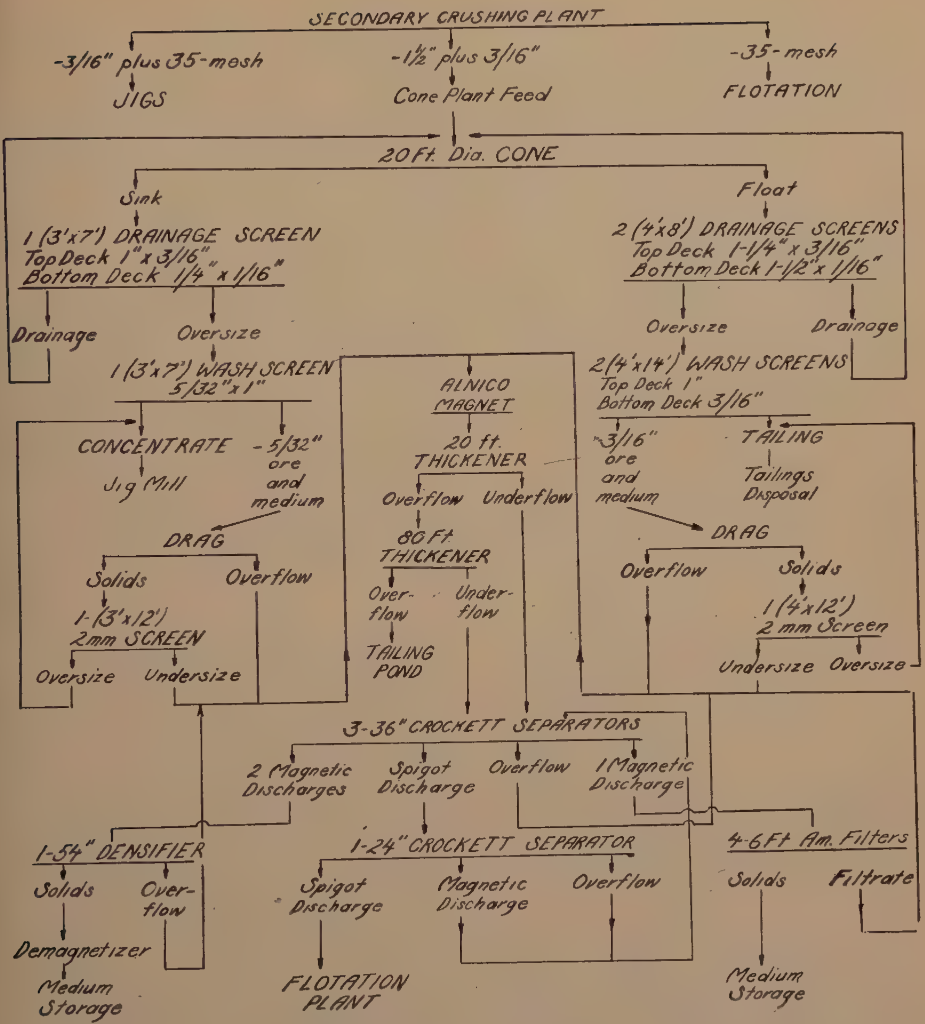

FIG 1—EAGLE-PICHER CENTRAL MILL. HEAVY-MEDIA SEPARATION PLANT.

CHARACTER OF MEDIA

The galena medium used was the normal lead flotation concentrate produced in the mill. This product was very fine, and tended to become more so through attrition as it was circulated within the heavy-medium plant and cleaning circuit. The settling rate of this material was very low. The circulating medium normally contained about 75 pct lead.

The ferrosilicon medium is a minus 100-mesh dry-ground product containing 15 and 85 pct of silicon and iron, respectively. The screen analysis of the material is:

MESH	PER CENT
−80 + 100	1.6
−100 + 200	26.7
−200 + 325	20.7
−325	51.0

Combined turbimetric and screen analyses indicate the material normally to have a median radius of 31.0 microns. The apparent specific gravity of the material varies between 6.85 and 7.25. When the specific gravity of the pulp is 2.70, the apparent viscosity of fresh medium is 2.68 centipoises. The settling rate yields 5 pct clear water with dirty medium after 5 minutes; this rate increases to 24 pct clear water with relatively clean medium. Medium made up with fresh ferrosilicon settles to 30 pct clear water after 5 minutes, and shows 36 pct clear water at time of ultimate compression, which is reached in 8 minutes.

FERROSILICON LOSS

The loss of ferrosilicon is 0.77 lb per ton of mill feed. This loss is distributed approximately as shown in Table 1 when treating 12,000 tons of ore per day.

TABLE 1—*Loss of Ferrosilicon*

Place in Process	Pounds		Percentage of Total Loss
	Total	Per Ton	
Cone tailing	2,620	0.218	28.3
Cone concentrate	180	0.015	1.9
Loss in magnetic separation	3,780	0.314	40.8
Oxidation and unaccounted for	2,670	0.223	29.0
	9,250	0.770	100.0

Tests indicate that 0.36 pct of the total ferrosilicon is initially oxidized as indicated by solubility in ammonium citrate, and after one week's immersion in mill water in a quiescent state the soluble loss increases to 0.85 pct. In the cone, where attrition occurs, this rate of oxidation is probably higher.

OPERATING CHARACTERISTICS

The most noticeable difference in the use of the two types of medium is in relative segregations due to the difference in their settling rates. The fine galena, for the period of operation considered, gave more stable pulps than the ferrosilicon. The average differential in specific gravities between the pulps at the top and bottom of the cone when using galena was only 0.03, whereas the differential when employing ferrosilicon averaged 0.15. The specific gravity at the top of the cone for both media was 2.69, but the bottom specific gravities were 2.72 and 2.84, respectively.

As will be noted later, this higher differential has caused a materially increased grade of cone concentrate. When the character of ore treated has been very clean, however, the differential has tended to increase so that excessive outside agitating air was required in order to prevent the cone from dropping its load, and choking the airlift. The necessity for this excessive air has led to increased losses in the tailing.

GRADE OF CONE PRODUCTS

The average cone tailing produced during the last 25 months of the galena operation contained 0.53 pct Zn. This compares with a tailing of 0.57 pct Zn for the first 19 months using ferrosilicon. A considerable period of time was required, however, to stabilize the ferrosilicon operation. The potential possibilities of the use of ferrosilicon at this plant have only recently become apparent. The tailing for August 1946, for example, contained 0.55 pct and for September and October, each 0.51 pct Zn.

The average grade of cone concentrate produced using galena was 11.05 pct Zn compared with an average of 16.00 pct for the ferrosilicon medium. This has resulted in material savings in grinding and treatment costs in the subsequent treatment of concentrate. When using galena, four No. 86 Marcy ball mills were required to regrind cone concentrate; with the new medium, three mills are sufficient and occasionally two mills are enough.

GRADE OF FLOTATION CONCENTRATE

One of the primary requirements at the Central mill is the production of zinc flotation concentrate with low lead content for the manufacture of pigment. This requirement was difficult to meet when galena was used as the medium. The galena circulated throughout the cone and the medium-cleaning plant; the medium-cleaning-plant tailing was returned to the head of the main flotation plant, and a portion of the lead concentrate was again returned to the cone circuit. This circulation of galena periodically built up a sulphated condition on a portion of the galena, which, being less floatable, finally reported in the zinc concentrate. This condition was one of the principal reasons for changing from galena to ferrosilicon. The grades of the flotation concentrates for the two periods (Table 2) indicate an average reduction of 0.25 pct in the lead content of the zinc concentrates.

TABLE 2—*Grades of Flotation Concentrates*

Medium	Zinc, Pct	Lead, Pct
Zinc concentrate, galena medium....	60.71	0.65
Zinc concentrate, ferrosilicon medium	60.80	0.40

CONSUMPTION OF MEDIUM

It was estimated that the galena loss was 0.6 lb per ton of cone feed whereas the corresponding consumption of ferrosilicon to Aug. 1, 1946 was 0.77. This average consumption of ferrosilicon has recently been reduced by the addition of a third 36-in. Crockett separator. The approximate value of galena at the present time is $120 per ton; the cost of ferrosilicon delivered at the mill is $88.74 per ton, so that the costs of the two media per ton of ore treated are very close.

COSTS OF CONE OPERATION

Comparative costs of the cone operation per ton of mill feed for the two periods are presented in Table 3. These costs do not include secondary crushing or preparation of cone-plant feed. The increased cost using the ferrosilicon is partly due to general increase in wages and cost of supplies rather than to any additional labor or additional supplies required.

TABLE 3—*Comparative Costs*

Item	Galena Medium	Ferrosilicon Medium
Labor..........................	$0.0145	$0.0175
Repairs and supplies..........	0.0150	0.0280
Power..........................	0.0095	0.0100
Medium.........................	0.0360	0.0340
Total cost................	$0.0750	$0.0895

COSTS OF BALL-MILLING

Costs of ball-milling have been reduced because the higher grade of cone concentrate means reduction in tonnage to grind. Comparative costs in terms of cost per ton of original mill feed are given in Table 4.

TABLE 4—*Costs of Ball-milling*

Item	Galena Medium	Ferrosilicon Medium
Labor..........................	$0.003	$0.003
Balls..........................	.017	.012
Power..........................	.013	.009
Liners.........................	.008	.005
Classifiers....................	.001	.001
Miscellaneous..................	.001	.001
Total..................	$0.043	$0.031

CONCLUSIONS

Metallurgical—The average net results for the two different types of medium indicate a higher grade of concentrate but a higher tailing loss using ferrosilicon.

Results obtained in the past three months on ferrosilicon, however, indicate that in the future a lower tailing and a higher grade of concentrate using ferrosilicon may be expected.

Costs—The cost of the operation exclusive of royalty when using galena was $0.075 per ton as compared with $0.0895 per ton with ferrosilicon. The galena operation had a tonnage of 507 tons per hour, against 482 tons per hour with the ferrosilicon medium, so that the costs as given are approximately equal.

Power—The average connected load with galena was 730 hp as compared with a present load of 711 hp. Average power consumptions for the two media were 0.746 and 0.954 kw-hr per ton, respectively. At the present time the power consumption compares favorably with that required using galena.

General—In general it is believed that the potentialities of the use of ferrosilicon are just beginning to be realized. Although the tailing loss for the first 19 months operation with ferrosilicon was higher than that of the last two years of the galena operation, for the past three months the tailing loss has equaled the best that was produced using galena.

The chief virtue of ferrosilicon is the ease of cleaning; it is believed that with some of the muddy ore now being received at the mill, it would be extremely difficult, if not impossible, to operate with galena.

The main difficulty with the use of ferrosilicon has been to keep the bottom gravity from becoming too high during the treatment of clean ore. When this condition has arisen, excess agitating air has been necessary, with consequent increase in tailing loss.

Summary

The results of the comparison described herein indicate that the use of ferrosilicon has been worth while, particularly considering the results attained during the past few months. The respective operating costs using the two media and the grades of tailings produced are approximately equal. With ferrosilicon the grade of cone concentrate is materially higher; the cost of fine-grinding the cone concentrate for flotation treatment is less; the lead content of the flotation zinc concentrate has been reduced; better treatment of muddy ores is possible because of the greater ease of cleaning the ferrosilicon medium.

A New Separating Vessel for Sink-float Concentration

By E. C. Bitzer,* Member AIME

(New York Meeting, March 1947)

The primary object of the work described in the following pages was to simplify the equipment in the separating circuit of the heavy-media process by substituting a spiral classifier for the separatory cone. Mechanical considerations were the leading attraction and the best that was hoped for on metallurgical performance was that the new vessel would give results equal to those obtained by conventional equipment.

Tests have demonstrated in commercial operation on iron ore that the mechanical advantages of the spiral separator are about as predicted; also, the metallurgical performance is superior to the cone on the ores treated.

Treatment of iron ore involves the separation of two products: a concentrate that sinks and a float product that contains true float material plus a controlled division of interfering middling. The middling material is heavier than the average specific gravity of the separating medium.

Recent work with the spiral separator on Tri-State zinc ore showed that the middling material had to be recovered as concentrate because this fraction contained the bulk of the zinc in the ore. It has been demonstrated that it is possible to make three products with the new separator on an ore such as this: a float tailing, a float middling, and a high-grade heavy sink product. It appears from this last work that the new separator has definite mechanical and metallurgical advantages over any type of separating cone for treating feed that contains large amounts of material of a specific gravity close to that of the separating medium. However, either type of separator should perform equally well, metallurgically, on feed that is free of middlings.

Each series of experiments with the spiral separator has required additional work to fully develop the possibilities of the machine. For this reason the paper is divided into chronological parts.

History of Development

A drag-type classifier was first proposed as a substitute for the cone separator by Butler Brothers technicians during the initial tests with galena medium at Crosby, Minn., in 1937. The desire for finding a substitute for the cone arose from trouble with interfering middlings and certain operating disadvantages of the cone. The possibility of using a spiral classifier was also suggested at this time by Emmett Butler and C. J. Abrams. However, inability to satisfactorily recover fine galena, which was used as a medium, terminated all development work at Crosby.

During the winter of 1937–1938 investigation started on magnetic medium with concurrent development of the inverted or closed-top cone, which was found necessary to eliminate iron-ore middlings with the float product. About

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* Colorado Iron Works Co., Denver, Colorado.

March 1938, a small-scale experiment was made with a 16-in. Akins classifier as a substitute for the cone but the results were indifferent and the idea was abandoned. All this work was conducted at the Mines Experiment Station of the University of Minnesota.

Following the development of ferrosilicon as a medium, the process was tried again commercially on the Mesabi Range during the ore season of 1939. Of all the plant equipment in use at that time the separatory cone gave the least trouble and during this period more suitable means for handling medium in the circuit were worked out. For several seasons, therefore, attention was directed toward the improvement of auxiliary equipment and, in view of the generally satisfactory operation of the cone, no further thought was given to a separatory vessel for plus ¼-in. ore.

An Esperanza drag classifier was used in Butler Brothers' experiments on minus ¼-in. iron ore at the Mines Experiment Station in 1939 but for a number of reasons not pertinent to the present discussion the project was abandoned.

During the ore season of 1943 an experiment was conducted with a 16-in. submerged-spiral Akins classifier at the Trout Lake concentrator of the Oliver Iron Mining Co. near Coleraine, Minn. Fine hematite was used as a medium and it has been reported that a concentrate was produced from a low-grade feed.

In 1942 the staff of the Colorado Iron Works Co. reviewed the possibilities of the Akins classifier as a separatory vessel and some preliminary inspections were made in operating plants using both the heavy-media and H. and H. concentration processes. Because of the pressure of war-begotten projects, however, it was impossible to do any actual testing until the summer of 1944.

By this time there was considerable incentive for finding a substitute for the closed-top cone on the Mesabi Range. The worst problems of the heavy-media flowsheet had been solved and the disadvantages of the cone were by contrast more apparent than they had been. Another factor that favored the experiment was that the ores available for treatment were becoming more refractory, with increasing amounts of middlings, and the metallurgical performance of the cone was not so good as it had been previously. Accordingly it was decided to revive the idea of trying an Akins classifier as a separating vessel and a machine was installed in Butler Brothers' Harrison washing plant near Cooley, Minnesota.

Pilot Test at Harrison Plant

FLOWSHEET

A 24-in. Akins classifier was installed in the plant as shown in the flowsheet (Fig 1). The primary object was to determine whether ferrosilicon medium could be maintained in suspension in the classifier tank and, if this were accomplished, to conduct a test under parallel conditions with the cone separator of the circuit in which the classifier was installed.

The flowsheet as shown was, in theory at least, parallel to the cone so far as medium circulation was concerned. However, the Wilfley pump for the classifier circulation medium was somewhat oversize and control of the discharge was irregular. In order to prevent choking of the pump-discharge line, it was necessary to circulate more medium than was required for overflowing float from the classifier pool. This fact probably accounted in part for the loss of fine sink material in the subsequent tests. Had time permitted, a by-pass for medium could have been installed but the operating season was well advanced toward freezing weather at that time, and it was decided to forego this and certain other obvious improvements to the testing arrangement

PRELIMINARY TESTS

One of the principal arguments advanced against trying a spiral classifier as a sink-float separating vessel was that a modification of the machine was already adapted to settle and dewater cleaned ferrosilicon medium. However, this argument was proved to have no foundation because the circulating plant medium showed little tendency to settle in the classifier and a separation of sink and float took place as soon as crude ore was fed to the machine.

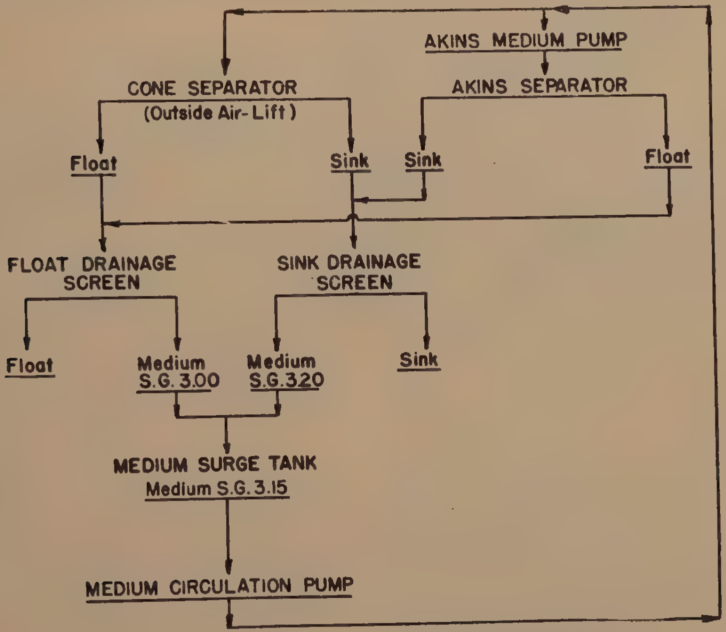


FIG 1—FLOWSHEET OF PILOT TEST AT HARRISON PLANT.

Several other characteristics of the machine remained to be established and the following is an excellent illustration of the fact that astute reasoning and theories cannot substitute for experimentation. During the first few days of testing a baffle was installed in the pool in various ways and the ore and medium were fed to the pool between the baffle and the overflow weir. The assumption was that the baffle would prevent float material from traveling forward with the movement of the spiral and thus prevent contamination of the sink product. The baffle was effective in retaining float on the surface but a certain amount of float sank with the heavy concentrate, owing to entrapment. This material eventually floated to the surface between the baffle and the point where the sink product emerged from the pool, and a heavy blanket of float built up in this area. However, instead of contaminating the sink product, the float actually forced itself under the baffle and finally reported in the overflow. Needless to say, as soon as this phenomenon was checked by assay, the baffle was removed and never used thenceforth.

The float material does not report with sink in the spiral separator because the specific gravity at the shallow end of the pool is too high to allow light material to sink. A considerable amount of this high-specific-gravity medium likewise drains out of the bed of sink as it is conveyed out of the tank, and the current caused by this drainage sweeps float material backward.

Compression of the concentrate bed, together with the presence of high-specific-gravity medium, also causes elimination of middlings from the sink product.

Also apparent from the preliminary work was the fact that, although a certain amount of float will be entrapped when the proportion of sink material is high, the light material will float if it is allowed enough time to free itself. The movement of the spiral was slow enough to permit this, and the sink product or concentrate from the spiral separator was exceptionally free from misplaced float material all through the subsequent tests.

After the discovery that any sort of baffle was superfluous, the experimental separator was operated as in ordinary classifier practice; that is, with the feed at the side and overflow at the back of the tank. The arrangement used for the remainder of the tests is sketched in Fig 2.

COMPARATIVE TESTS WITH CONE SEPARATOR

Ore Feed for Spiral Separator

The spiral separator was installed below a dewatering screen that delivered minus ¾-in., plus ¼-in. washed feed to the cone. A plate with a hinged connection extended into the screen discharge chute so that varying amounts of cone feed could be diverted to the separator. The cone feed rate amounted to about 100 gross tons per hour and between 7½ and 14 tons was taken for Akins feed. The ore traveled by gravity in a steel chute to the separator tank and circulating plant medium was added to the ore about 3 ft ahead of the point where it entered the pool.

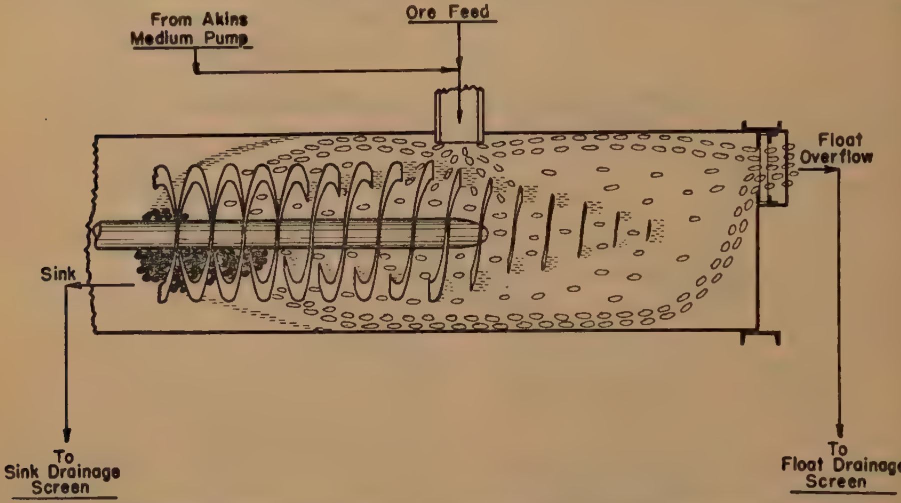


FIG 2—ARRANGEMENT OF FEED AND PRODUCT DISCHARGE OF 24-INCH SEPARATOR DURING TEST AT HARRISON PLANT 1944.

Method of Sampling

Samples of spiral separator feed, tailing, and concentrate were taken at regular intervals, which varied between 15 and 30 min, depending upon how long the machine was operated. A 30-min interval would permit operation for 8 hr before the sample containers were filled. Corresponding samples were taken of the cone products at the same time. Because of the uncertainty involved in splitting this relatively coarse material, the individual sample cuts were not divided in the plant and sampling was discontinued when all sample pails were filled.

Specific Gravity of Separation

The specific gravity of circulating medium for both spiral and cone was identical. From test to test it varied from 2.85 to 3.25.

Laboratory Tests

All samples were split into two equal parts in the laboratory. Half was then crushed and prepared for assay. The other half was reserved for laboratory sink-float and screen tests for determining efficiency of plant operation. The sink-float tests were made with ferrosilicon medium at the average gravity at which the plant had been operated for the sampling period. Because of knowledge of segregation in the feed samples, no sink-float tests were made on this material.

Test Results

Table 1 shows the average results of nine days of comparative sampling between the spiral and the cone.

TABLE 1—*Average Results of Nine Days Sampling*

Comparison of Sink-float Efficiency Tests, Weighted Averages

Product	Spiral			Cone		
	Pct Wt	Fe	SiO_2	Pct Wt	Fe	SiO_2
Conc (sink).	94.24	61.76	6.20	85.60	61.28	6.82
Conc (float).	5.76	51.49	18.88	14.40	44.42	27.13
	100.00	61.17	6.93	100.00	58.85	9.74
Tails (sink).	44.83	54.30	13.63	42.81	55.61	12.62
Tails (float).	55.17	36.66	39.59	57.19	33.23	44.83
	100.00	44.57	27.95	100.00	42.81	31.04

Comparison of Product Assays

Method	Feed		Concentrate		Tailings	
	Fe	SiO_2	Fe	SiO_2	Fe	SiO_2
Spiral........	55.51	14.50	61.95	6.40	45.26	27.30
Cone.........	55.80	14.36	59.66	9.12	44.76	28.65

Comparison Based on Screen-test Assays

Method	Feed		Concentrate		Tailings	
	Fe	SiO_2	Fe	SiO_2	Fe	SiO_2
Spiral........	55.02	14.24	61.17	6.93	44.57	27.95
Cone.........	55.19	14.58	58.95	9.74	42.81	31.04

Metallurgical Results

All comparisons gave about the same story. The efficiency of the spiral separator on the concentrate was a decided advantage and it was felt that the operating conditions did not give the machine a fair chance of producing a tailing as low as might have been possible. It was obvious from the test data that a lower specific gravity in the spiral separator would have produced a lower tailing with the same concentrate assay as the cone.

Mechanical Comparisons

Elimination of the airlift and the ability to go through power interruptions without draining the separator tank were all decidedly in favor of the spiral separator. Also, inasmuch as the amount of medium removed by the spiral was only 10 pct of that discharged by the airlift for the same tonnage of concentrate, an additional economy in medium pumping was to be expected when the new separator was substituted for the cone.

These advantages were important factors in operation, and in view of the reasonable explanations for the slightly higher tailing from the new separator it was decided to experiment with a machine of commercial size during the next ore season.

Comparative Test between 78-inch Spiral Separator and 7½-Foot Closed-top Cone

The results of the test with the 24-in. Akins separator were very encouraging but not sufficiently conclusive to warrant the design of a heavy-media plant of commercial size and to gamble the production of a season's operation upon the probable performance of large units. Therefore, it was decided to design and install a single 78-in. machine in the plant along with cone separators of conventional Mesabi construction.

Flowsheet of Spiral Separator Circuit

The heavy-media section of Butler Brothers' Patrick washing plant during the 1945 season consisted of two 7½-ft closed-top cone separators with separate drainage and washing screens, pumps, medium-cleaning equipment, and other accessories. Each unit could operate independently.

A 78-in. separator was installed as close as possible to No. 2 cone unit so that either the spiral or the cone could be used with the same auxiliary equipment.

The chutes following the feed-preparation screens were arranged so that the ore feed could be split into coarse and fine fractions and either size could be delivered to either No. 1 or No. 2 cone.

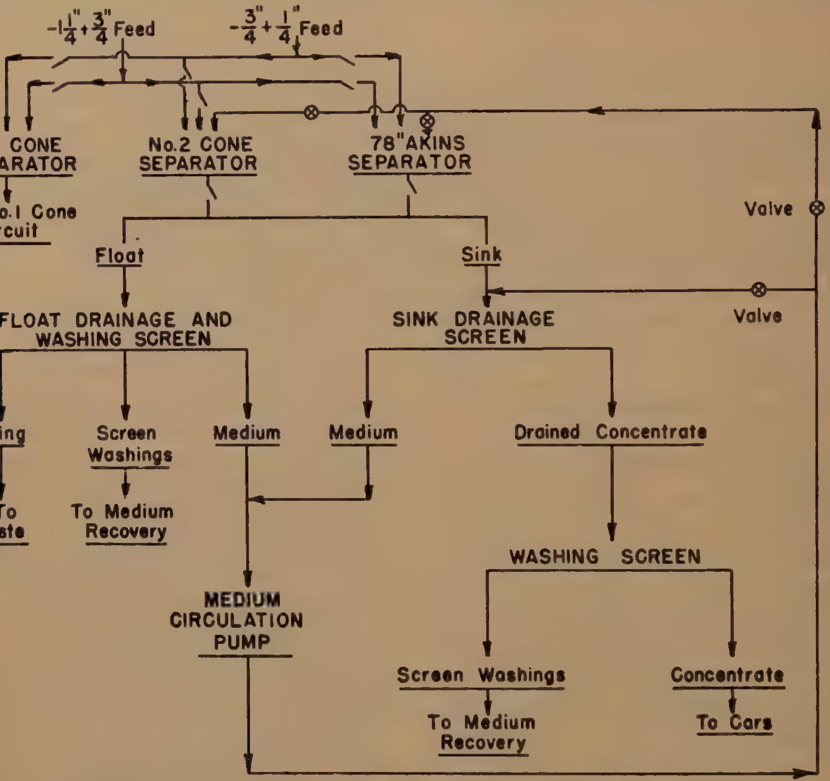


Fig 3—Flowsheet of 78-inch separator circuit, Patrick plant, 1945.

Fig 3 illustrates the flexibility of the flowsheet.

A portion of the circulating medium in the spiral separator circuit was diverted to the sink-product discharge launder. The reason for this was that the slope of the launder was too flat for gravity flow to the drainage screen and the extra medium was used for sluicing. Had it not been necessary to do this the power consumption for the pump for circulation of medium would have been appreciably less for the spiral separator than for the cone.

Preliminary Testing

When the spiral separator was put into operation, No. 2 cone was shut down. At the start no attempt was made to segregate the feed ore into fine and coarse fractions and both the spiral and No. 1 cone received minus 1¼-in. plus ¼-in. feed. The total tonnage for heavy-media treatment was about 375 gross tons per hour, of which the spiral received 200 tons and the cone 175 tons. The discrepancy was unintentional and was caused by uneven distribution of ore on the feed-preparation screens. This also led to differences in the grade of feed to the two circuits, which is apparent in the comparative assays given in the next section.

SAMPLING AND LABORATORY PROCEDURE

Procedures similar to those employed for the pilot test of the preceding season were followed for this test. This time, however, it was possible to operate the spiral separator without the cone and the laboratory sink-float tests were made at specific gravities corresponding to plant operation of both separators.

TEST RESULTS

At the end of one week of continuous operation the average results shown in Table 2 were obtained.

TABLE 2—*Results of One Weeks Operation with Spiral Separator*

Product	Assays, Pct			
	Spiral		Cone	
	Fe	SiO_2	Fe	SiO_2
Feed	52.08	18.65	54.40	15.67
Concentrate	57.38	10.65	56.35	12.29
Tailings	42.63	32.85	43.23	29.96

The results showed the performance of the new separator to be superior but the discrepancy in the character of feed to the two circuits was too great to make definite conclusions. Moreover, observation of the action of currents in the pool of the spiral separator indicated that better results might be obtained if the feed and overflow points were reversed.

In accordance with the practice followed in the small-scale operation of the previous season, the separator had been arranged to receive feed at the side of the tank with the tailings overflow at the lower end of the pool. Obviously with this arrangement slow-settling sink material or fines would be carried a considerable distance toward the tailing overflow before it would sink sufficiently to be caught by the spiral and raked forward. Inasmuch as the direction of travel of the fines had to be thus reversed, the time of their retention in the pool was longer than necessary and the chances for the material to report in the overflow were improved. Another factor that operated against recovery of the fines was that flow of medium *backward* to the weir against the currents caused by the spiral moving medium and sink material *forward* probably caused eddy currents that interfered with the recovery of fine sink material.

These considerations led to a decision to discontinue operation of the spiral separator to make alterations necessary to reverse the feed and tailing overflow launders. Accordingly the machine was shut down, the spiral was raised, and the medium was left in the tank. Because of a shortage of mechanics and welders, about 30 days elapsed before the alterations were completed.

FINAL COMPARATIVE TEST

The new feed and overflow arrangement of the spiral separator is shown in Fig 4. No other alterations were made to the machine, and after it was put into operation for the second time it was operated continuously for the remainder of the washing season—from about August 1 to October 15.

TEST CONDITIONS

The operation of the new separator and the control of specific gravity in both the spiral and No. 2 cone were left entirely in the hands of the plant crew.

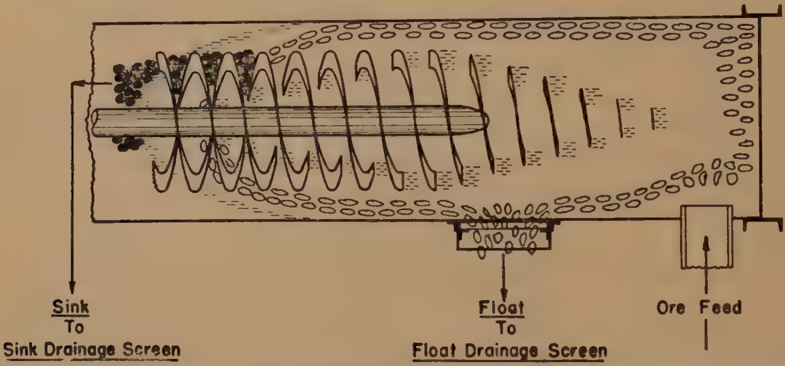


FIG 4—ARRANGEMENT OF FEED AND PRODUCT DISCHARGE OF 78-INCH SEPARATOR DURING TEST AT PATRICK PLANT, 1945.

Sampling for metallurgical comparison was generally done on the day shift, although on several occasions some sampling was done at other times to check the effect of certain plant conditions.

The arrangement of the feed distributor ahead of the separators permitted operation of the spiral separator on minus ¾-in., plus ¼-in. feed one week and on minus 1¼-in., plus ¾-in. feed the following week; thereby, comparison of spiral results for alternate weeks with those of the cone gave a reasonably accurate comparison of the performance of both types of separators. It is necessary to qualify the preceding statement because ore was delivered from three pits during the test and absolutely parallel test conditions were never achieved.

DISCUSSION OF RESULTS ON SIZED FEED

It is apparent from Table 3 that the cone was almost as effective as the spiral in producing a clean concentrate from the coarse fraction of feed. Laboratory checks showed 6.13 pct float in the cone concentrate as against 5.34 pct float in the spiral concentrate. This is as it should be because coarse particles should separate rapidly in any kind of vessel. The spiral tailing was decidedly better, however.

The laboratory tests also showed why the silica content of the cone concentrate was lower than that in the spiral product. A

TABLE 3—*Comparison of Operation on Sized Feed*

Kevin ore, minus 1¼-inch plus ¾-inch (four tests)

Product	Spiral Separator, Medium Sp Gr 3.04, Average			Cone Separator, Medium Sp Gr 3.06, Average		
	Pct Wt	Fe, Pct	SiO_2, Pct	Pct Wt	Fe, Pct	SiO_2, Pct
Feed.......	100.00	52.02	19.76	100.00	51.98	18.75
Concentrate.	80.25	55.88	13.30	70.01	56.94	11.30
Tailing.....	19.75	36.34	43.04	29.99	40.40	36.08
Iron recovery.......	86.20			76.69		

Laboratory Check Separations

Product	Pct Wt Sink	Fe, Pct	SiO_2, Pct	Pct Wt Float	Fe, Pct	SiO_2, Pct
Spiral concentrate..	94.66	56.44	12.36	5.34	45.86	30.03
Spiral tailing	13.34	49.87	22.59	86.66	34.36	46.19
Cone concentrate..	93.87	57.83	9.87	6.13	43.23	33.14
Cone tailing.	41.03	53.28	16.39	58.97	31.43	49.78

perfect separation with the spiral would have yielded only 12.36 pct silica whereas the cone would have produced a 9.87 pct silica with 100 pct separation of sink and float. This is an illustration of the practical impossibility of obtaining parallel feed characteristics over any considerable period of time.

The comparison of results on the fine size range, Table 4, is clearly in favor of the spiral separator in every respect. Of particular interest in this comparison is the character of the sink fraction of both the spiral and cone tailings. Both tailings contained about 30 pct of sink material but the sink in the spiral tailing ran 20.51 pct silica as against 12.91 pct silica in the sink fraction of the cone tailing. The sink fraction of the spiral tailing is high in iron (51.30 pct) but the silica content prohibits inclusion of the material in the concentrate. Material such as this influences Mesabi heavy-media practice to an extent not generally understood by outsiders, and the subject will be discussed in more detail later.

TABLE 4—*Comparison of Operation on Sized Feed*

Kevin ore, minus ¾-inch, plus ¼-inch (four tests)

Product	Spiral Separator, Medium Sp Gr 2.88, Average			Cone Separator, Medium Sp Gr 3.04, Average		
	Pct Wt	Fe, Pct	SiO_2, Pct	Pct Wt	Fe, Pct	SiO_2, Pct
Feed......	100.00	54.67	14.75	100.00	53.56	17.73
Concentrate.	82.29	57.54	10.92	72.38	56.46	13.06
Tailing.....	17.71	41.33	34.50	27.62	45.96	29.29
Iron recovery.......	86.7			76.3		

Laboratory Check Separations

Product	Wt, Sink	Fe, Pct	SiO_2, Pct	Pct Wt, Float	Fe, Pct	SiO_2, Pct
Spiral concentrate..	94.35	57.93	10.38	5.65	51.02	19.93
Spiral tailing	30.04	51.30	20.51	69.96	37.04	40.50
Cone concentrate..	81.59	58.22	10.23	18.41	48.67	25.60
Cone tailing.	30.84	56.72	12.91	69.16	41.17	36.60

GENERAL COMPARISON OF RESULTS

The proceeding comparison was made on ore from one source only. A final comparison was calculated from the results of 13 tests on three different ores with the spiral (and the cone) treating both size ranges (Table 5).

TABLE 5—*All Ores, Both Size Ranges* (13 *Tests*)

Product	Spiral Separator, Medium Sp Gr 2.93, Average			Cone Separator, Medium Sp Gr 3.07, Average		
	Pct Wt	Fe, Pct	SiO_2, Pct	Pct Wt	Fe, Pct	SiO_2, Pct
Feed.......	100.00	54.23	16.20	100.00	53.70	16.99
Concentrate.	82.93	57.32	11.83	75.69	56.59	12.60
Tailing.....	17.07	39.22	38.46	24.31	44.70	30.49
Iron recovery.......	87.66			79.76		

Laboratory Check Separations

Product	Pct Wt Sink	Fe, Pct	SiO_2, Pct	Pct Wt Float	Fe, Pct	SiO_2, Pct
Spiral concentrate..	95.64	57.72	11.17	4.36	48.54	26.38
Spiral tailing	27.06	50.15	22.76	72.94	35.16	44.28
Cone concentrate..	87.87	57.87	10.55	12.13	47.34	27.41
Cone tailing.	35.54	55.30	14.30	64.46	38.85	39.41

DISCUSSION OF GENERAL RESULTS

The results in Table 5 follow the same general pattern of those on the minus ¾-in., plus ¼-in. feed tabulated in Table 4 and verify the metallurgical superiority of the spiral separator.

As shown in the laboratory check separations, the sink fraction of the spiral tailing runs lower in iron and higher in silica than the corresponding fraction of the cone tailing. The weight also is appreciably less, which indicates that the cone was rejecting merchantable concentrate with the tailings.

MECHANICAL CHARACTERISTICS OF THE SPIRAL SEPARATOR

Power Consumption

The power required to operate the spiral varied with the size of ore feed. About 15 hp input was recorded for a full load (200 tons per hour) of minus 1¼-in. plus ¾-in. concentrate. When the finer size range, minus ¾-in., plus ¼-in., was treated the power input dropped to as low as 8 hp. This was due partly to a lower tonnage of the finer size, which was approximately 150 tons per hour. Fine material likewise offered less resistance to raking than coarse concentrate.

The net saving in power in the spiral-separator circuit was the difference between the power required for the cone drive and air compressors and that required for the drive of the spiral. Power input to the cone drive and compressors was not determined but obviously the approximate saving was represented by the elimination of compressed air from one side of the plant.

Feed Surges

Because of rapid changes in structure of the ore received at the plant, the amount of plus ¼-in. material recovered for heavy-media treatment varied over a wide range and caused surges of cone and spiral feed. The cone was quite sensitive to an increase in feed rate and would choke unless the feed was shut off quickly following a surge. It was especially vulnerable in this respect when treating the coarse fraction of feed.

The spiral was not affected by surges, and for this reason the operators preferred to treat coarse feed on the spiral side of the plant as much as possible, because the practice resulted in continuity of feed and more production.

Shutdowns

The tank of the spiral separator was not drained for the entire season. Every week the plant was shut down 24 hr for general repairs, during which period it was necessary either to drain No. 1 cone or operate the air compressor and cone drive to maintain circulation of medium.

On one occasion, noted previously, the spiral was shut down for 30 days with medium in the tank and the machine was started without difficulty.

Circulation of Medium

Less circulating medium was required to remove products from the spiral separator than from the cone. Normally this would have been reflected in decreased power consumption of the medium-circulation pump. As explained previously, in connection with the flowsheet Fig 3, the pump had to furnish extra medium to sluice the spiral concentrate to the drainage screen. Moreover, the separator was set higher than would have been necessary had the cone not been installed on that side of the plant. This imposed an additional head on the medium pump and a few feet of extra head on this line makes a considerable difference in power requirement.

METALLURGICAL CHARACTERISTICS OF THE SPIRAL SEPARATOR

Concentrate Grade

The difference in grade between spiral and cone concentrate was due partly to a lower percentage of misplaced material (float) in the spiral product. This in turn is caused by the fact that concentrate remains longer in the spiral separator than in the cone and the additional residence time provides the opportunity for entrapped float particles to free themselves from the concentrate load on the spiral.

Specific Gravity Differential in the Pool

Generally speaking, the specific gravity in the spiral-separator pool increases from the point of feed entrance at the deep end of the pool to the shore line, where the sink product emerges. The specific gravity is also slightly higher on the side where the spiral is rotating upward than on the opposite side. There is no appreciable differential between top and bottom *when the medium is stabilized.*

Elimination of Middling Material

One reason for the superiority of the spiral separator for iron-ore concentration was that the machine eliminated more high-silica middling material in the overflow. Because of the higher specific gravity of the pool near the shore line, it was impossible for middling particles to sink at this point, and they built up in the pool until they eliminated themselves in the overflow by mechanical crowding. This explains the anomaly of float particles having a higher specific gravity than the medium with which they overflow.

This characteristic of the machine was fortuitous and, frankly, the basic reason was not understood during the period of testing with iron ore.

Mesabi Heavy-media Metallurgy

Concentration of Mesabi ores by heavy media has generally been regarded as being a crude adaptation of the process because of the general high iron content of the tailings. Actually, some very fine work has been accomplished in view of the character of the material being treated and the rigid conditions imposed by the economics of marketing iron ore. Consequently, the following discussion is in no sense an apology for the preceding metallurgical data. Many millions of tons of merchantable concentrate would still be unshipped were it not for the adaptation of the heavy-media process to the problem.

Character of Ore

The average ore being treated by heavy media in Minnesota does not conform to the usual conception of sink-float feed; it is not all blue hematite and white silica. The production from some pits has aptly been described as consisting of nothing but poor concentrate and rich middling. In addition, most deposits contain large quantities of slimy material that cannot be completely removed prior to heavy-media treatment. This material fouls the medium and affects the action of the separating vessel. Slime also imposes unusual burdens on the medium-cleaning and recovery equipment.

Design of Cone Separator

The presence of large amounts of middling material in the average ore deposit was responsible for the unusual design of the separating cone originally used for iron ore. The upper part of the cone is constricted in the form of an inverted cone to cause a rising current that is effective in preventing the accumulation of middling material in the vessel. In an ordinary open cone this material will concentrate in the vessel until a point is reached at which it must settle. It will settle with disastrous results to the airlift unless controlled. The constricted (closed-top) cone was effective in preventing the accumulation of middling material but the rising current also interfered with the separation of fine sink material, thus limiting the process to about 4-mesh bottom size.

Adaptation of Akins Classifier

As stated at the beginning of the paper, the original purpose of the trial of the Akins classifier was to demonstrate some mechanical ideas. The metallurgical results were fortuitous because the action of the spiral in conjunction with differentials of specific gravity within the pool could

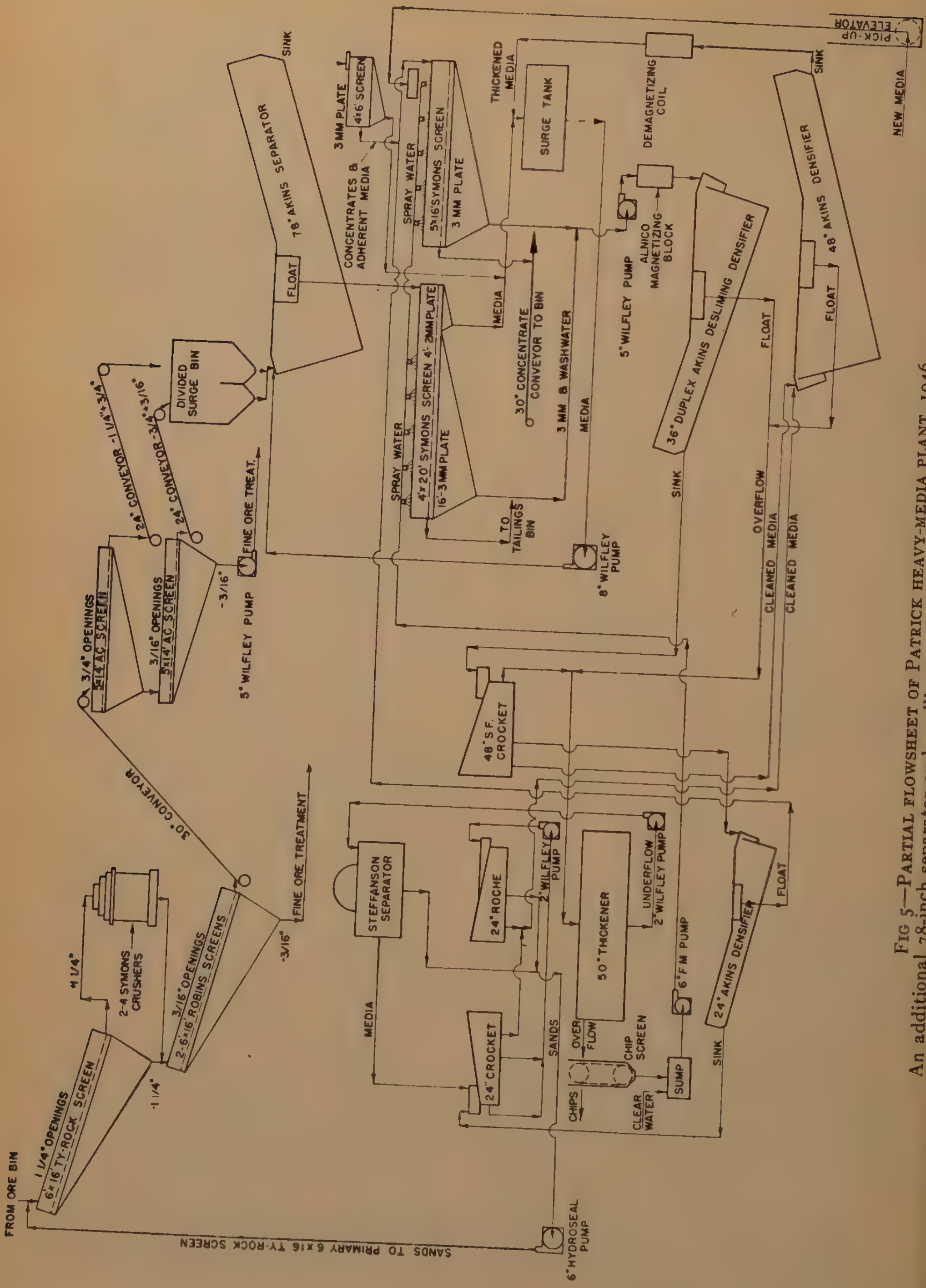


FIG 5.—PARTIAL FLOWSHEET OF PATRICK HEAVY-MEDIA PLANT, 1946.
An additional 78-inch separator and auxiliary equipment constitutes remainder of equipment.

not have been predicted in advance of experimentation.

New Installations

Two plants have been converted to the use of the Akins separator—the Patrick plant, operated by Butler Brothers near Cooley, and the Hill-Trumbull plant, operated by the Mesabi Cliffs Mining Co., between Marble and Calumet, Minn. Another plant is now being built, by the Stanley Mining Co., near Biwabik, for operation in 1947.

Flowsheet of New Patrick Plant

Fig 5 is a partial flowsheet of the Patrick plant as it was operated during the 1946 washing season. In addition to the replacement of the cones by spiral separators, the conventional thickener ahead of the magnetic separators was replaced by special densifiers, and the arrangement of the magnetic separators is novel.

Operation of the plant has been successful but a description of the operating features and metallurgy would require a separate paper.

Test on Lead-zinc Ore

Interest in the work with the new separating vessel on the Mesabi Range led to another pilot test, with a 24-in. machine, at the Central mill of the Eagle-Picher Mining and Smelting Co., at Cardin, Okla. An investigation of heavy-media tailings at the Central Mill had indicated that a possible reduction of 0.20 pct zinc might be made in the tailing loss if a separation were made at a specific gravity of 2.65. The cone separator had seldom been able to operate at a specific gravity of less than 2.70.

The work with the spiral separator on this test led to modifications in the operation of the machine that make possible its application to a much broader field of sink-float separation than had been indicated by operation on iron ore. The results of the work also clarified several characteristics of the machine that before had not been completely understood. One of the most important of these was the mechanics of middling separation in the pool, and it was demonstrated that it is possible to reject middlings separately from heavy concentrate and light tailings.

FLOWSHEET OF TEST AT CENTRAL MILL

The cone separator at the Central mill is 20 ft in diameter with open-top design and an "inside" airlift; that is, the concentrate is discharged above the center of the cone and the underflow or "sink" medium returns to the vessel by gravity. By contrast, the cones on the Mesabi Range have "outside" airlifts and the sink medium is mixed with the overflow medium and returned to the cone by a pump. The distinction was a fundamental one in the test to be described and will be better understood by a comparison of Figs 1 and 6. At the Central mill the only medium available for operating the spiral separator was that from the cone overflow, which is the lowest in specific gravity. This is shown by the typical medium gravities in Fig 6.

MEDIUM CIRCUIT

A 4-in. Wilfley pump delivered part of the cone float drainage medium to a 24-in. Akins separator through a junction box. The box had an overflow that was designed to maintain a constant head on the discharge pipes leading to the experimental separator. These pipes were valve-controlled as indicated in Fig 6. All these features were advantages over the arrangement at the Harrison plant (see Fig 1).

ORE FEED TO SPIRAL SEPARATOR

Heavy-media feed was delivered to the cone at a rate of approximately 450 net tons per hour by a belt conveyor. A small steel bin was installed on the operating floor under the belt-conveyor head pulley and a swing chute was designed to cut a portion of the main ore stream and divert it into the bin. A belt conveyor delivered ore from the bin to the experimental separator and an adjustable gate on the bin was used to control the feed rate.

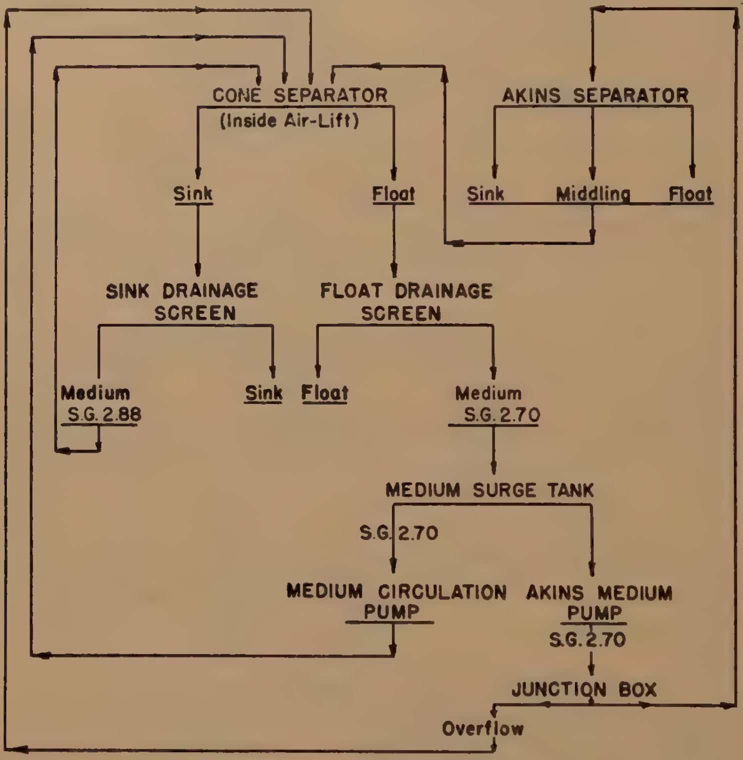


Fig 6—Flowsheet of pilot test at Central mill.

The bin had a live capacity of about one ton and it was necessary to cut into the main stream about every 10 min. to maintain constant feed to the spiral separator. The small bin capacity was advantageous because frequent filling ensured a representative sample of plant feed and eliminated uncertainties that might have arisen from frequent changes in the type of ore feed.

Sampling Procedure

Samples of spiral-separator feed and products were taken every 30 min. When it was desired to check any product by laboratory sink-float tests two cuts were taken and placed in separate sample containers.

The wash screens for cone tailings were double-decked and the washed tailings were separated into fine and coarse fractions. On this account it was not possible to take hand samples of cone tailings at the same time spiral-separator samples were being taken, as had been done in previous testing. The plant tailings were sampled automatically and were collected at the end of each shift. Therefore the results obtained with the spiral were not strictly comparative because it was not always possible to operate the machine for a full shift.

HEAVY-LIQUID TESTS

The investigation of the cone tailings prior to the test with the spiral separator was conducted by checking daily samples with acetylene tetrabromide. Sink-float separations were made at 2.65 and 2.70 sp gr, and the fraction recovered between these points represented the reduction that could be expected in tailing loss if a separator could be efficiently operated at a sp gr of 2.65. Samples of spiral-separator products were checked in the same manner and at the same specific gravities, so the results could be correlated with those from the previous investigation.

Therefore the results of the heavy-liquid tests made on both spiral and cone products in this instance indicate only how much material of critical gravity was rejected by the respective separators. Had the products been checked at the actual operating gravities, which was about 2.60 for the spiral and 2.70 for the cone, the results would have indicated the efficiency of each separator.

PRELIMINARY TESTING

It had been previously demonstrated on the Mesabi Range that the spiral separator could not recover middling material. This was an advantage in treating iron ore but was fatal to successful separation on this particular lead-zinc ore.

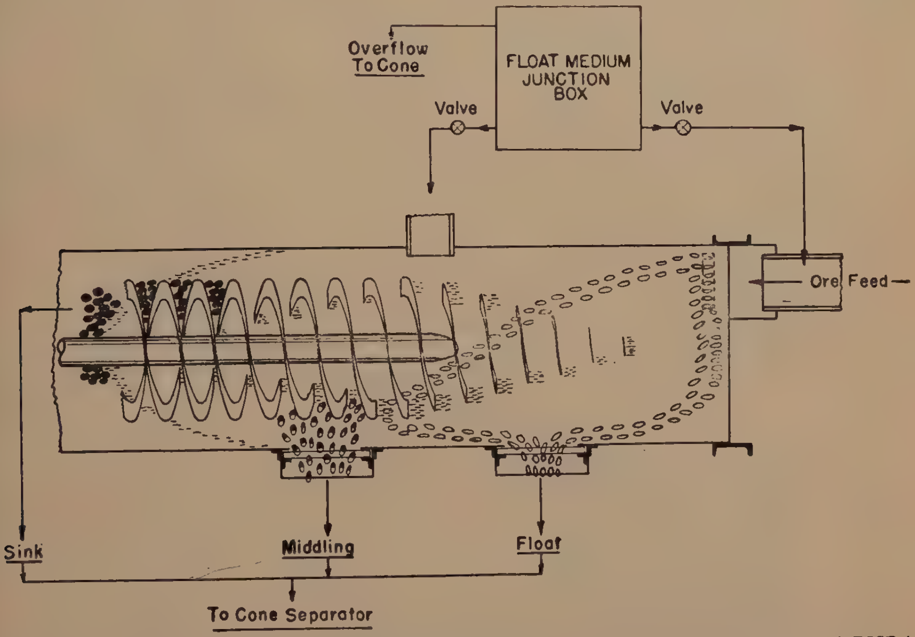
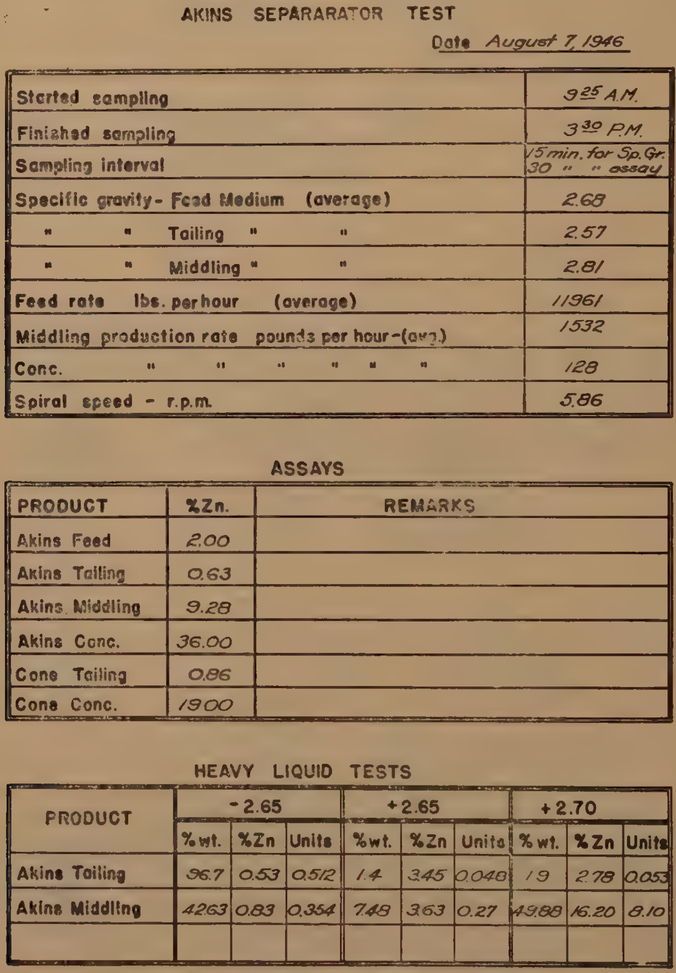


FIG 7—ARRANGEMENT OF FEED AND PRODUCT DISCHARGE OF 24-INCH SEPARATOR DURING TEST AT CENTRAL MILL, 1946.

Examination of the pool indicated that middlings congregated at the shallow end near the shoreline and an overflow opening was cut into the tank at this point, to allow them to discharge. As soon as this was done the middlings ceased to appear in the tailings. This was the first indication that a three-product separation was feasible.

It was soon learned, however, that a considerable quantity of float material was being carried forward on the surface of the pool, and there was nothing to prevent this material from overflowing with middlings. This difficulty was remedied by introducing medium to the surface of the pool on the side opposite the tailing and middling overflows and about midway between these points. A current was thereby created which prevented tramp float material from reporting in the middling overflow. Fig 7 shows an approximate plan arrangement of the spiral separator as described and Table 6 shows the assays of middling products before and after the introduction of extra medium at the side of the tank.



AKINS SEPARARATOR TEST

Date *August 7, 1946*

Started sampling	9^{25} A.M.
Finished sampling	3^{30} P.M.
Sampling interval	15 min. for Sp. Gr. 30 " " assay
Specific gravity - Feed Medium (average)	2.68
" " Tailing " "	2.57
" " Middling " "	2.81
Feed rate lbs. per hour (average)	11961
Middling production rate pounds per hour-(avg.)	1532
Conc. " " " " " "	128
Spiral speed - r.p.m.	5.86

ASSAYS

PRODUCT	%Zn.	REMARKS
Akins Feed	2.00	
Akins Tailing	0.63	
Akins Middling	9.28	
Akins Conc.	36.00	
Cone Tailing	0.86	
Cone Conc.	19.00	

HEAVY LIQUID TESTS

PRODUCT	- 2.65			+ 2.65			+ 2.70		
	%wt.	%Zn	Units	%wt.	%Zn	Units	%wt.	%Zn	Units
Akins Tailing	96.7	0.53	0.512	1.4	3.45	0.048	1.9	2.78	0.053
Akins Middling	42.63	0.83	0.354	7.48	3.63	0.27	49.88	16.20	8.10

Fig 8—Typical test on lead zinc ore.

Tests to Determine Concentration Ratio

The preliminary work established the method of operating the spiral separator to recover a separate middling product and a series of tests was made to determine the concentration ratio. On account of the spotty nature of the ore it was necessary to determine the ratio by actual weights. Fig 8 is reproduction of the record of a typical day's operation, which will illustrate the method of obtaining the various data.

Tables 7 and 8 are condensed summaries of a series of tests and Table 9 shows the distribution of zinc losses in the tailing.

TABLE 6—*Results of Preliminary Tests*

Test No.	Medium, Sp Gr		Assays, Pct Zn			
	Feed	Mid-dlings	Feed	Tail-ings	Mid-dlings	Con-cen-trates
6/21	2.56		2.28	0.47	2.71	34.50
6/22	2.53	2.78	1.70	0.56	2.85	37.80
6/24	2.50	2.74	1.58	0.47	2.18	28.00
6/27[a]			1.90	0.58	6.96	30.40

[a] Cross current used for first time.

DISCUSSION OF RESULTS

Control of Specific Gravity

The specific gravity of the feed medium to the spiral separator was generally the same as that of the cone overflow, although in a few tests some water was added to the spiral pump sump to experiment with lower gravities.

TABLE 9—*Spiral Tailings, Individual Assays*

Screen Size, In.	−2.65 Sp Gr		−2.70 = 2.65 Sp Gr		+2.70 Sp Gr	
	Pct Total Wt	Zn, Pct	Pct Total Wt	Zn, Pct	Pct Total Wt	Zn, Pct
+7/8	38.18	0.36				
+3/8	49.33	0.36	0.30	1.63	0.11	2.05
−3/8	11.61	0.22	0.11	2.15	0.36	2.75

Spiral Tailings, Calculated Composite Assays

Screen Size, In.	Pct Wt	Pct Zn	Units
+7/8	38.18	0.36	0.137
+3/8	49.74	0.37	0.185
−3/8	12.08	0.31	0.038
			0.360

TABLE 7—*Tests to Determine Concentrate Grade and Recovery*

Date	Sp Gr, Medium			Spiral Product Assays, Pct Zn					Cone Assays, Pct Zn	
	Feed	Tails	Middling	Feed	Tails	Midd.	Conc.	Comb. Conc.	Tails	Conc.
6/28	2.65	2.52	2.78	1.85	0.47	5.96	36.00	8.95	0.68	17.3
6/29	2.67	2.53	2.81	2.06	0.34	4.06	36.80	6.39	0.63	16.2
7/1	2.70	2.57	2.83	1.90	0.52	4.51	32.90	5.82	0.72	16.2
7/2	2.70	2.58	2.82	1.60	0.58	6.01	28.00	6.98	0.61	16.2
7/3	2.70	2.58	2.87	1.10	0.59	6.33	27.20	7.28	0.67	19.5
8/7	2.68	2.57	2.81	2.00	0.63	9.28	36.00	10.73	0.86	19.0
Avg	2.68	2.56	2.82							

Combined spiral concentrate (weighted average) = 7.72 pct Zn.
Concentration ratio (calculated from time samples) = 6.15 to 1.
Percentage of recovered zinc in middling = 68.8 pct.
Percentage of recovered zinc in sink conc. = 31.2 pct.
Average (arithmetical) spiral tailing = 0.52 pct Zn.
Average (arithmetical) cone tailing = 0.69 pct Zn.
Average (arithmetical) cone concentrates = 17.6 pct Zn.

TABLE 8—*Heavy-liquid Tests on Spiral Products*

Date	Product	−2.65 Sp Gr			−2.70 +2.65 Sp Gr			+2.70 Sp Gr			Total Zn
		Pct Wt	Zn, Pct	Units	Pct Wt	Zn, Pct	Units	Pct Wt	Zn, Pct	Units	Units
6/28	Tailings	95.6	0.47	0.44	2.2	5.00	0.11	2.2	4.55	0.10	0.65
6/29	Tailings	95.1	0.26	0.25	4.2	1.73	0.07	0.7	3.95	0.03	0.35
7/1	Tailings	98.1	0.50	0.49	1.9	2.46	0.046				0.536
7/2	Tailings	96.7	0.58	0.56	2.6	2.48	0.06	0.7	2.40	0.01	0.63
7/3	Tailings	96.9	0.56	0.54	2.2	2.26	0.06	0.9	2.95	0.02	0.62
8/7	Tailings	96.7	0.53	0.51	1.4	3.45	0.048	1.9	2.78	0.053	0.611
8/7	Middlings	42.63	0.83	0.354	7.48	3.63	0.27	49.88	16.20	8.10	8.724

It was soon learned that the valve controls on the medium lines from the junction box to the experimental separator were of little use, because if the volume of circulating medium in the pool was too low the specific gravity of the tailing overflow would drop to the point where too much barren gangue would begin to report with middling material. The reason for this was that the middling overflow was in effect robbing an already impoverished medium. The specific gravity at the middling discharge averaged 0.26 more than at the tailing overflow. Incidentally, the spiral removed no medium from the separator except that attached to the sink material. The amount of medium discharged with sink product, therefore, was insignificant.

The average specific gravity of tailing medium was 2.56 and the actual separation of combined concentrate and tailings was probably made at close to 2.60. The excessive volume of medium flowing through the separator pool unquestionably interfered to some extent with the specific-gravity separation because the amount of plus 2.65 sp gr material in the tailings averaged 3.5 pct.

These conditions amount to a departure from a true specific-gravity separation, which can be closely approached with the machine if the conditions are right. The feed medium must be high enough in specific gravity to permit balancing the gravities between the tailing and middling overflows without the necessity of resorting to velocity to control the settling rate of the medium.

Elimination of Middling

The method of removing middling from the spiral separator is a combination of mechanical action bv the spiral and simple gravity overflow. Since nothing can stop the discharge of middlings, they will not accumulate in the pool and cause trouble. These are advantages that the cone does not have. Erratic middling elimination was the fundamental difficulty with the cone on this particular application because the quantity of middling material in the plant feed varies over a considerable range.

General Conclusions

A definite concentration and a low tailing were achieved with the spiral separator. The fact that the tailing averaged lower than the plant tailings indicates that an improvement in tailing loss may be realized by a separation at a lower gravity. This verifies the conclusions reached from the original investigation.

The low grade of concentrate produced was a result of excessively low specific-gravity separation, and there was no practical means for correcting this condition.

Two distinct grades of concentrate were produced, a middling that floated and a heavy sink of high grade. Since both products had to be ground to flotation size for final concentration, there was no advantage to be gained by recovering the products separately. The possibility of such a separation is interesting, however. For instance, a separation of fluorspar might be made in which the tailing could be rejected, the middling sent to fine grinding and flotation, and heavy concentrate marketed directly.

Acknowledgments

The results summarized herein represent about three years of investigation and the work of many individuals, some unnamed here, who contributed to the success of the development of the separator.

Acknowledgment is made to the management of Butler Brothers, St. Paul, Minn., and that of the Eagle-Picher Mining and Smelting Co., Joplin, Mo., for permission to publish the results of plant testing and operation.

Special acknowledgment is made to the assistance of A. E. Matson and L. J. Erck,

research engineer and assistant research engineer, respectively, for Butler Brothers at Cooley, Minn. Mr. Erck, by his suggestions and interest in the problem, was largely responsible for the early successes in the undertaking.

The author's thanks are also due to E. H. Crabtree, Jr., chief metallurgist, and Walter Freudenberg, mill superintendent, of the Eagle-Picher Mining and Smelting Co., for their whole-hearted cooperation in the tests at the Central mill. Heine Kenworthy, of the Rolla Station of the U. S. Bureau of Mines also participated in this work, and his help and suggestions were invaluable.

Finally, acknowledgment is gratefully made to the staff of the Ore Dressing Laboratory of the American Cyanamid Co. at Stamford, Conn., for valuable suggestions and help throughout the period of this investigation.

DISCUSSION

E. H. Crabtree, Jr.*—The test at the Central Mill using the spiral separator indicated that this unit could be operated safely at much lower specific gravities than could the cone, and for this reason the separator could make lower tailings than could the cone, at a sacrifice of grade of concentrate and middling products. A specific gravity of circulating medium of below 2.65 is essential to improve the cone tailings over their normal run. At such a gravity the cone is operating at the danger point, while the separator is able to operate at gravities considerably below this point without mechanical difficulty. The fact that the separator could not make lower tailings than could the cone and at the same time maintain a high enough grade of combined concentrates to keep from loading up the subsequent concentrate retreatment plant was an economic rather than a metallurgical problem; in planning a new plant this fact could be taken into consideration.

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* Eagle-Picher Mining and Smelting Co., Joplin, Mo.

A Suggested Approach to the Analysis of Mineral Suspensions by High-frequency Electrical Measurements

By John D. Morgan, Jr.,* Junior Member, and Sylvain J. Pirson,† Member AIME

An instantaneous and continuous analysis of a mineral suspension should be of great value in controlling various mineral preparation processes. Described herein is a method of analysis based on the use of high-frequency alternating current, which, under certain restrictive conditions, could be of use.

Theoretical Considerations

In a suspension of conductive minerals, the electrical effect primarily responsible for the variations in electric conductivity is the capacitance. In Fig 1*a*, let an alternating voltage E be applied across two electrodes F and F', between which are two rows (S and T) of mineral particles, all of approximately the same size, immersed in a fluid. Let the number of rows of particles between the electrodes be R and the number of particles in each row be N. If A is the area of a particle, D is the distance separating two particles, and K is a constant depending on the fluid, the capacity, C_0, of the particle system will be given by:

$$C_0 = \frac{KRA}{ND}$$

Referring now to Fig 1*b*, let R, the number of rows, be held constant, but let N, the number of particles in each row (S and T), be increased X-fold. The capacity, C_1, of

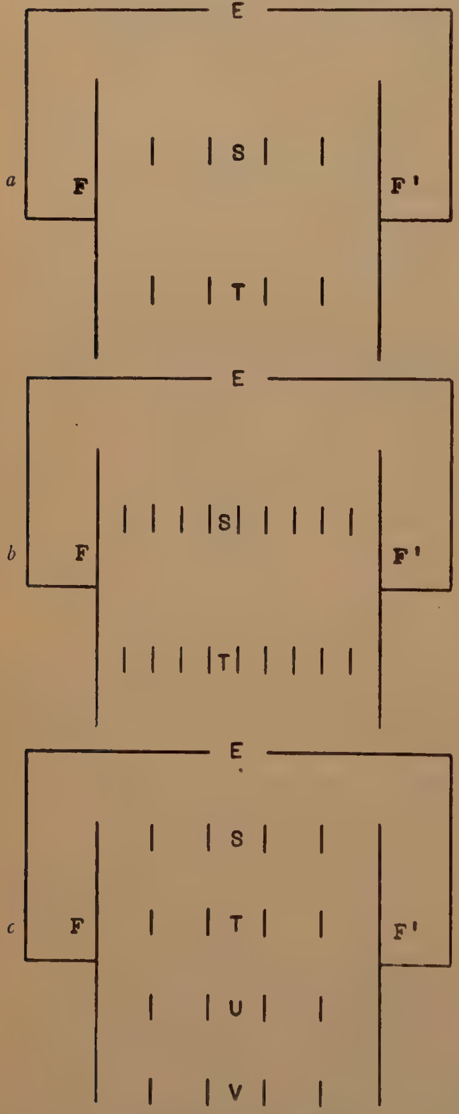

Fig 1—Illustration of theory of high-frequency analyzer.

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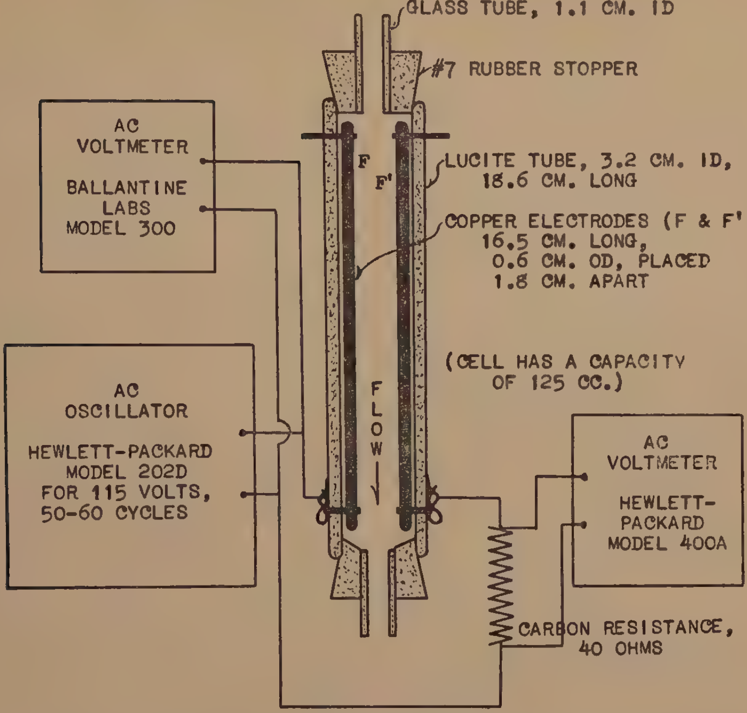
GLASS TUBE, 1.1 CM. ID
#7 RUBBER STOPPER
AC VOLTMETER
BALLANTINE LABS MODEL 300
F
F'
LUCITE TUBE, 3.2 CM. ID, 18.6 CM. LONG
COPPER ELECTRODES (F & F'), 16.5 CM. LONG, 0.6 CM. OD, PLACED 1.8 CM. APART
(CELL HAS A CAPACITY OF 125 CC.)
AC OSCILLATOR
HEWLETT-PACKARD MODEL 202D FOR 115 VOLTS, 50-60 CYCLES
FLOW
AC VOLTMETER
HEWLETT-PACKARD MODEL 400A
CARBON RESISTANCE, 40 OHMS

FIG 2—HIGH-FREQUENCY ANALYZER.

the system remains unchanged as may be seen from:

$$C_1 = \frac{KRA}{XN\frac{D}{X}} = C_0$$

However, if R, the number of rows, is increased Y-fold (to become S, T, U, V, as illustrated by Fig 1*c*), all other things being held constant, the new capacity, C_2, of the system becomes:

$$C_2 = \frac{KYRA}{ND} = C_0Y$$

Since a general increase in the number of particles in circulation between the electrodes F and F' will increase both R and N, an increase in the capacity of the system can be expected. Consequently, causing more particles to pass between the electrodes should cause an increase in the electrical current flowing between them, since for a purely capacitive circuit:

$$I = E2\pi FC$$

where I is the current flowing in the circuit, E is the impressed voltage, F is the frequency, and C is the capacity. The ohmic resistance of the particle suspension will of course also play some part in determining the amount of current that will pass between the two electrodes. Indeed, in the case of nonconductive minerals, the electrical effect primarily responsible for the variations obtained is the ohmic resistance, for here the nonconductive particles present in suspension reduce the conductivity of the more conductive liquid in which the particles are suspended.

HIGH-FREQUENCY SUSPENSION ANALYZER

Fig 2 is a schematic diagram of the essential elements of the analyzer. All of the electronic equipment illustrated operates on a line voltage of 115 v, 60 cycles. The oscillator generates a signal of variable voltage and frequency: for the tests de-

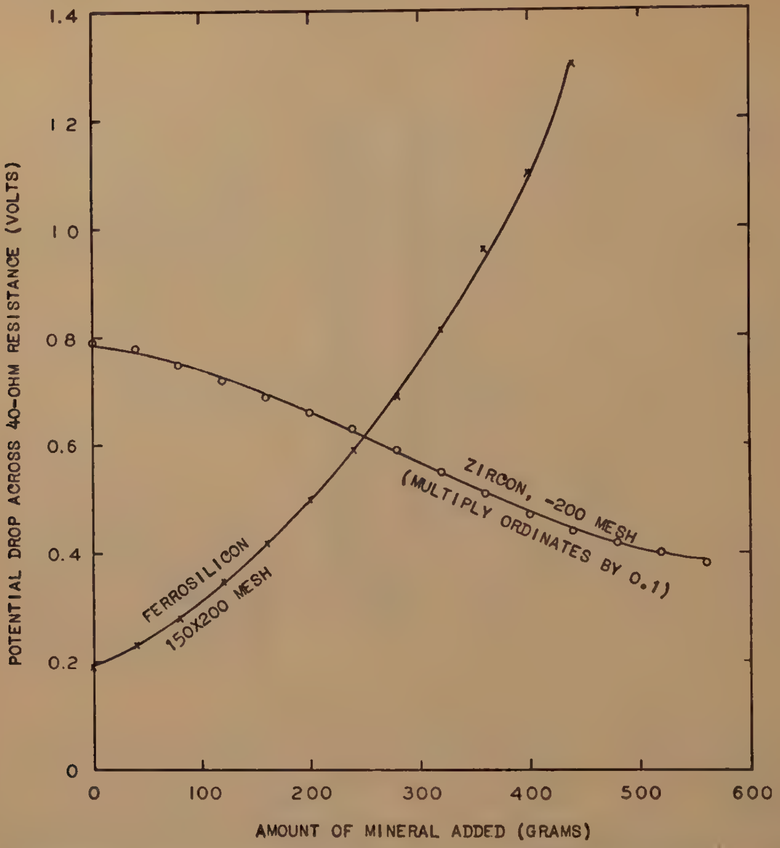

Fig 3—Contrasting behavior of a conductive and a nonconductive mineral in high-frequency analyzer.

scribed herein (in which the amount of mineral in suspension was varied) the input voltage to the analyzer circuit was held constant at 4.0 v, and the input frequency was held constant at 50,000 cycles. These values were determined by experimentation to be most satisfactory. One voltmeter was employed to monitor the input voltage to the analyzer, while the second voltmeter was used to measure the potential drop across a resistance of known value placed in series with the analyzer cell. A suspension was circulated through the cell (in the direction indicated by an arrow on the sketch) by an air lift in the glass tubing liquid circuit in series with the cell. The cell was kept in a vertical position for all tests. The total volume of the cell plus the circulating system was about 800 cc, while the volume of the cell alone was 125 cc.

Liquid Part of Suspension

The water employed in making all the tests (except the water-quartz-galena suspension test) was tap water which had been allowed to stand in contact with the mineral used for a period of at least 12 hr to eliminate changes of liquid conductivity due to later solution of electrolytes. The water was then decanted and the mineral carefully dried so that the latter could be added in weighed increments. To obtain reproducible results, the concentration of electrolyte in the liquid must be both small and constant, for it was found that 0.1 pct by

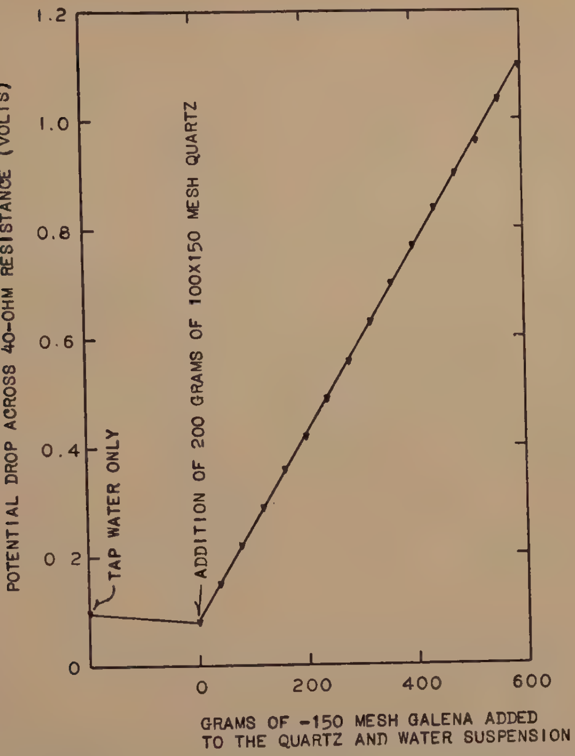

Fig 4—Behavior of a conductive mineral when added to a nonconductive mineral in high-frequency analyzer.

weight of salt (NaCl) would produce a conductivity change equivalent to that produced by the maximum possible circulating load of most conductive minerals. In each test an initial liquid load of 500 cc was circulated in the cell, and to this liquid the suspensoid was added in 40-g increments. The suspension was then allowed to circulate through the cell until equilibrium conditions were obtained, whereupon, ensuring that the input voltage was constant at 4.0 v, the potential drop across the standard 40-ohm resistance was read.

Results of Various Tests Where Amount of Mineral in Suspension Was Varied

Table 1 summarizes the results obtained in selected tests with the representative minerals, quartz, zircon, galena, and ferrosilicon. In these tests the amount of mineral present in the suspension was the variable.

Fig 3 contrasts the behavior of a typical conductive and a typical nonconductive mineral.

Fig 4 illustrates the curve obtained when a conductive mineral is added to a nonconductive mineral already in suspension. (In this test plain tap water was used, together with well-washed quartz and well-washed galena that had been air-dried.)

Fig 5 is based on the same data as the previous curves but has calculated densities as abcissas.

Result of a Test When Frequency Was Varied

Table 2 summarizes the results obtained from a test in which 100 cc of galena sludge (−150 mesh) in 500 cc of water which had

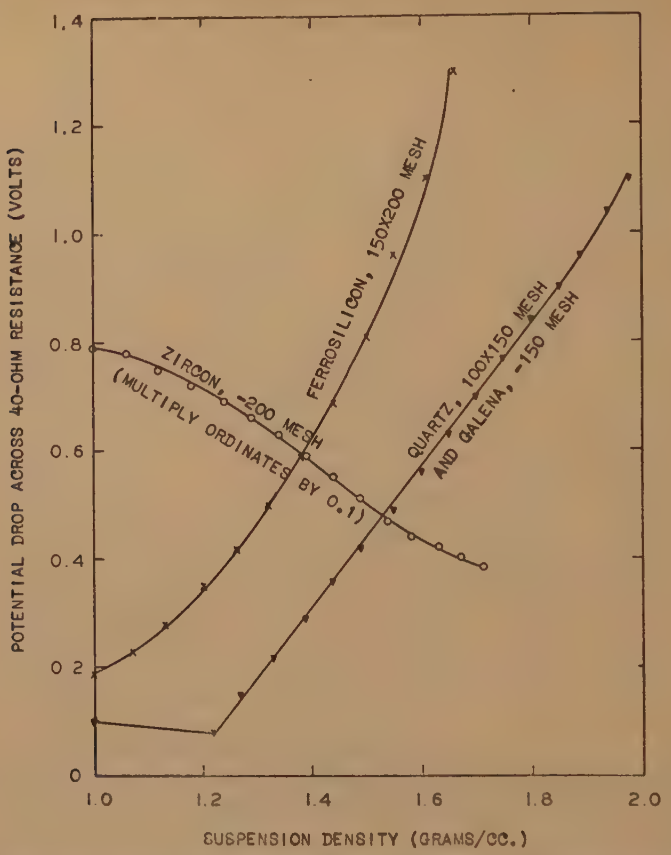

FIG 5—HIGH-FREQUENCY ANALYZER APPLIED TO MEASUREMENT OF DENSITY OF MINERAL SUSPENSIONS.

TABLE 1—*Summary of Representative Tests Using the High-frequency Suspension Analyzer*[a]

Total Grams of Mineral Added to 500 cc of Tap Water	Ferrosilicon, 150 × 200 Mesh		Zircon, −200 Mesh		Galena,[b] −150 Mesh	
	Calculated Suspension Density (g per cc)	Potential Drop Across 40 Ohm Resistance (Volts)	Calculated Suspension Density (g per cc)	Potential Drop Across 40 Ohm Resistance (Volts)	Calculated Suspension Density (g per cc)	Potential Drop Across 40-Ohm Resistance (Volts)
0	1.00	0.19	1.00	0.079	1.22	0.08
40	1.07	0.23	1.06	0.078	1.27	0.15
80	1.13	0.28	1.12	0.075	1.33	0.22
120	1.20	0.35	1.18	0.072	1.39	0.29
160	1.26	0.42	1.24	0.069	1.44	0.36
200	1.32	0.50	1.29	0.066	1.49	0.42
240	1.38	0.59	1.34	0.063	1.55	0.49
280	1.44	0.69	1.39	0.059	1.60	0.56
320	1.50	0.81	1.44	0.055	1.65	0.63
360	1.55	0.96	1.49	0.051	1.70	0.70
400	1.61	1.10	1.54	0.047	1.75	0.77
440	1.66	1.30	1.58	0.044	1.80	0.84
480			1.63	0.042	1.85	0.90
520			1.67	0.040	1.89	0.96
560			1.71	0.038	1.94	1.04
600					1.98	1.10

[a] All tests made with input voltage of 4.0 v, at 50,000 cycles.

[b] The galena was added to a suspension of 200 g of 100 × 150 mesh quartz. Before the quartz was added, the 500 cc of tap water alone had a density of 1.00 and a potential drop of 0.10 v.

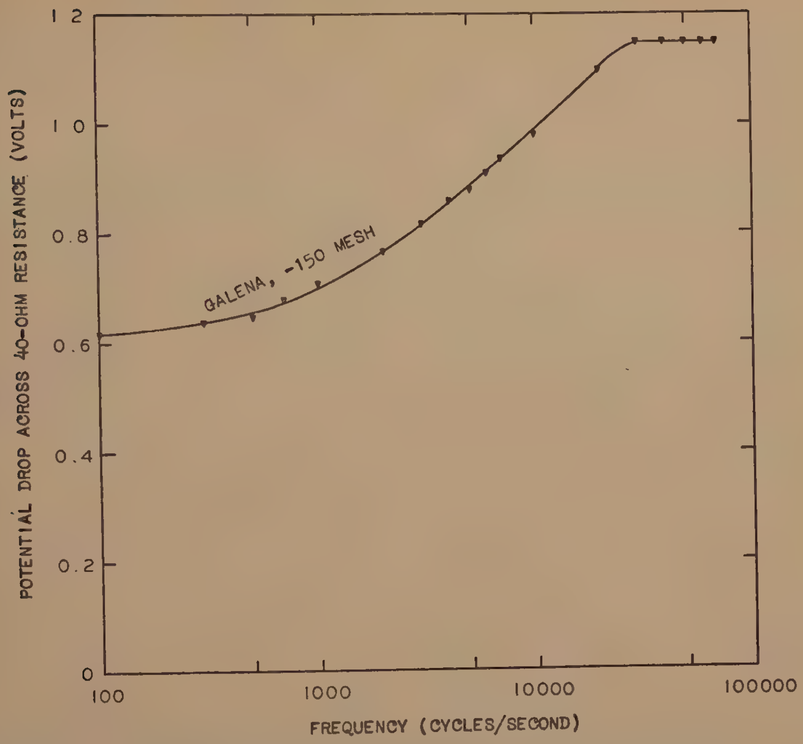

Fig 6—Effect of increasing frequency on conductive mineral suspension in high-frequency analyzer.

Table 2—*Effect of Frequency upon the Conductivity of a Mineral Suspension*

Water used: tap water allowed to stand overnight in contact with the galena. 500 cc used.

Galena used: −150 mesh material, 100 cc of wet sludge used.

Input voltage: 7.0 v.

Applied Frequency (Cycles Per Sec)	Potential Drop Across 40 Ohm Resistance (Volts)
100	0.62
300	0.64
500	0.65
700	0.68
1,000	0.71
2,000	0.77
3,000	0.82
4,000	0.86
5,000	0.88
6,000	0.91
7,000	0.94
7,000 (repeat at change of meter range)	0.94
10,000	0.98
20,000	1.10
30,000	1.15
40,000	1.15
50,000	1.15
60,000	1.15
70,000	1.15

been allowed to leach the galena overnight was circulated through the analyzer while the frequency was varied. The galena sludge was not air-dried in this test but was added wet.

Fig 6, based on the data of Table 2, definitely shows that there is an increase of conductivity at higher frequencies, thus validating the original premise that the current increase obtained was due primarily to the capacity effect.

Conclusions and Recommendations

It is realized that more tests of the method suggested should be made before definite conclusions as to its applicability are announced. However, since circumstances do not permit the authors to continue the investigation at the present time, the following conclusions are drawn from the work performed to date with the thought in mind that the information should prove useful in certain applications:

1. To analyze a suspension composed of a liquid of constant low conductivity and a single mineral of either higher or lower conductivity than the liquid, the method described could possibly find ready application.

2. To analyze a suspension composed of a liquid of constant low conductivity and a two-component mineral mixture in which one mineral is of low conductivity and one mineral of high conductivity, the method described could be applicable in determining only the amount of high-conductivity mineral present.

3. To determine the density of a mineral suspension, or to maintain such density between certain limits, and if such suspension is composed of a liquid of constant low conductivity and a single mineral of either lower or higher conductivity than the liquid, the method described could be again applicable.

Metallurgical Efficiency—a Yardstick in Lead-zinc Flotation Metallurgy

By R. A. Pallanch,* Member AIME

(New York Meeting, March 1947)

Probably one of the most perplexing problems with which a mill operator contends is the proper evaluation of his mill results. True, he accurately determines his recoveries of metals and grades of products monthly, weekly, and often daily. These figures do not of themselves directly translate to mill economics. When one figure is unusually good there may be an offsetting circumstance. The advantage of high lead or zinc concentrate grade may be offset by poorer recovery; high lead recovery may be offset by poorer zinc recovery, or vice versa; and so on. Besides, grades or recoveries may not of themselves mean more than a reflection of the grade and character of the ore during the period under study.

What the operator sorely needs is some one measuring stick that tells him when he has achieved his greatest return, regardless of grades or recoveries—for if he knows his methods and practices are yielding the optimum return he need no longer concern himself with the assays of products and tailings, or the corresponding distribution of metal values.

At first blush it would appear that the obvious answer to this is the actual money return to the mill. Paradoxical as it may seem, however, money return does not make a dependable yardstick any more than do grades and recoveries, as ore conditions would vary enough to bring about greater money return during periods when the actual metallurgical work is not as good. The mill operator, of course, should always aim for the greatest return on the ore, once it is in the mill bins, with proper attention always to mill treatment cost.

The need of a proper yardstick becomes more imperative with the increasing complexity of the problem. In the case of a simple, single-product operation on a fairly uniform ore, in which only one metal is of significant importance, the need is not great, as grade and recovery pretty well tell their own story. At the other extreme is the operation that makes three or four separate marketable products from a complex mixture of heterogeneous ores that vary tremendously in individual metal contents and ratios, degree of oxidation, and other characteristics, and in which gold, silver, copper, lead and zinc are all significant values.

Economic Recovery

The first attempt at obtaining a proper single yardstick for a fairly complex ore with which I am familiar is the use of the "Economic Recovery" term at the lead-zinc flotation mill of the Sunnyside Mining and Milling Co., Eureka, Colorado. This method had been applied there early in Sunnyside flotation-mill history (not later than 1920) and credit for the innovation I believe belongs to Messrs. Kuryla and Marquand. The term "Economic Re-

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* Mill Superintendent, U. S. Smelting, Refining and Mining Co., Midvale, Utah.

covery" as here defined was the percentage ratio of actual mill return per ton ore to the gross value per ton ore during any period under observation. It made a very convenient and satisfactory method of evaluating Sunnyside metallurgy during long periods when the ore remained fairly uniform in character and grade.

At the Midvale mill of the United States Smelting Refining and Mining Co., however, this method of evaluating mill work soon proved very misleading, particularly as the percentage of individual lease ores and custom ores in the mill feeds increased. It was soon evident that poor work on an ore that had a high degree of precious-metal value would show considerably higher Economic Recovery than would excellent work on an ore that had principally base-metal value.

METALLURGICAL EFFICIENCY

In order to retain the advantages of the method and largely rule out the disadvantages, we went a step further at Midvale (1936). We developed the term "Metallurgical Efficiency," which we defined as the percentage ratio of actual mill return per ton ore to the theoretical maximum return per ton ore if perfect grades and recoveries had been accomplished. We chose metallurgical perfection as the one base that could not be changed (or ever reached) no matter what the grades, recoveries, or degree of comminution. This term and method has been employed at the Midvale mill ever since.

ADJUSTED METALLURGICAL EFFICIENCY

One further step had to be taken, however. Because of the fact that Metallurgical Efficiency tends to increase with the grade of ore, we decided upon an adjustment, entirely empirical, to be applied to ores of varying grades. We assumed a set figure as a base standard of maximum theoretical return and adjusted to this standard. For instance, an ore that returns $8.00 from a theoretical maximum return of $10.00 figures 80 pct metallurgical efficiency. The same ore returning $26.00 from a theoretical maximum of $30.00 figures 86.6 pct metallurgical efficiency. (Such an extreme variation of a single ore classification is not unusual.) The difference in efficiency is due entirely to grade. If we take the base standard of $20.00 for theoretical maximum return for both grades of ore, and assume 90 pct efficiency for that portion of the value below or above the standard, we are able to arrive at an adjustment that tells the story much better. Thus, in the case of the lower grade ore the maximum return is $10.00 less than the base of $20.00; then 90 pct of $10.00, or $9.00, is added to the $8.00 return per ton, making $17.00. The adjusted metallurgical efficiency is $17.00 ÷ $20.00 = 85.0 pct.

In the case of the higher grade ore, the maximum return exceeds the base of $20.00 by $10.00. Applying 90 pct metallurgical efficiency to this $10.00 gives $9.00. Deducting this $9.00 from the $26.00 actual return makes a return of $17.00 for a $20.00 base. Again we have 85 pct adjusted metallurgical efficiency. In this way the grade effect is minimized and straight across comparisons may be made.

Having established a base theoretical maximum return per ton ore, we are then able to apply the adjusted metallurgical efficiency to each tonnage contained within a mixture and summate to what the adjusted efficiency for the mixture should be. If the mill results on the mixture show greater return, mill work has surpassed expectation; if less, mill work has fallen below expectation. Previous individual test work on the separate ores, figured to adjusted metallurgical efficiencies, with proper summation on the tonnage basis, is what establishes the expectation.

It would be absurd to imply that if another $10.00 of possible value were put into (or taken out of) an ore the metallurgical efficiency would be increased (or decreased) by the amount shown in the adjustment. Merely increasing the theoretical maximum of a partially oxidized ore, for instance, might not improve its efficiency, but actually might decrease it, because the metallurgical efficiency of the additional increment, if oxidized, would be much lower than the 90 pct figure used. The adjusted efficiency figure, however, does have the great advantage of making it possible to translate the work on individual ores to that on the combined tonnages within a mixture of the ores.

Illustration

By way of illustration, let us assume four separate ores of varying grades and proportions, which individually behave very differently as follows:

Ore	Percentage of Mixture	Gross Value	Maximum Return	Actual Return
A	15	$10.00	$ 6.00	$ 2.00
B	25	10.00	6.00	4.80
C	45	30.00	20.00	17.00
D	15	40.00	28.00	24.00
Summation	100	$23.50	$15.60	$12.75

Using 90 pct adjustment to $20.00 base, the following results are obtained:

Ore	Economic Recovery, Pct	Metallurgical Efficiency, Pct	Adjusted Efficiency, Pct
A	20.0	33.3	73.0
B	48.0	80.0	87.0
C	56.7	85.0	85.0
D	60.0	85.7	84.0
Mix	54.2	81.7	83.5

Applying the individual economic recoveries to the mixture direct gives a summated economic recovery of 49.5 pct, which does not check at all with the 54.2 pct in the table.

Applying the individual metallurgical efficiencies to the mixture direct gives a summated metallurgical efficiency of 76.1 pct, which does not check with the 81.7 pct in the table.

Applying the individual adjusted efficiencies to the mixture direct, however, gives a summated adjusted metallurgical efficiency of 83.5 pct, which exactly checks the figure in the table. This method of comparison is accurate.

The base of $20.00 per ton for theoretical maximum return may be changed to any other base, of course. Likewise the 90 pct adjustment may be changed to some other figure more in line with the circumstances. It is well to choose a base that is fairly average for mill ore conditions, and a percentage adjustment that is in line with the best mill work.

SOME SIMPLE METAL MARKET SCHEDULES

No reference has been made in the foregoing pages to any actual metal market or schedule, for that is local for each operator. However, it may be wise, by way of illustration, to assume some very simple, streamlined schedules, it being understood that any similarity to existing schedules and contracts is "purely coincidental."

Assume a schedule for lead concentrate as follows:

All of gold at $32.00
95 pct of silver at market
90 pct of copper at market less 6 cents
90 pct of lead less 1.5 pct off assay, at
  market less 1.5 cents
Iron pay, 6 cents per unit
Deductions, sulphur
  + treatment + freight.. $9.00 per dry
  ton

Assume a schedule for zinc concentrate as follows:

80 pct of gold at $35.00
80 pct of silver at market
Copper—no pay
80 pct of lead in excess of 3 pct at market
  less 2 cents
80 pct of zinc at market
Deductions, treatment
  + freight.......... $30.00 per dry ton

Assume a schedule for pyrite to a copper smelter as follows:

95 pct of gold at $35.00
95 pct of silver at market
50 pct of lead at 4 cents off price
Iron excess over silica, 5 cents per unit

Deduction, treatment plus freight, $6.00 per dry ton

Assume the war ceiling prices of:

Au	Ag	Cu	Pb	Zn
$35.00	$0.71	$0.12	$0.065	$0.0825

Metallurgical perfection would be achieved if all the copper mineral (assume chalcopyrite) and all the lead mineral (galena) were combined into a product that contained all the gold and silver, with no free pyrite, gangue, or other non-pay minerals; and all the zinc mineral were contained in a product made up of zinc mineral only. Any pyrite product then made would not pay its own way and therefore would have negative value, if recovered.

If a separate copper product is being marketed as such to a copper smelter, a copper-smelter schedule would apply in order to properly establish maximum pay for copper. This possibility is not being considered here, as thought is being given only to a plant making gold-silver-copper-bearing lead concentrate, gold-silver-lead-bearing zinc concentrate, and gold-silver-lead-bearing pyrite concentrate. Any copper existing in zinc and pyrite concentrate would represent a loss.

CALCULATION OF THEORETICAL MAXIMUM RETURN

A perfect galena concentrate would run 86 pct lead and 14 pct sulphur. Such a product would give a lead return as follows:

2000 × (0.860–0.015) × ($0.065–$0.015) × 0.9	$76.00
Total deduction	9.00
	$67.00

Thus, 1720 lb of lead returns a maximum of $67.00 = $0.039 per pound.

A perfect chalcopyrite concentrate would run about 34 pct copper and 30 pct iron. Shipped on the lead schedule gives:

2000 × 0.34 × 0.9 × ($0.12 – $0.06)	$36.70
Iron pay, 30 × $0.06	1.80
	$38.50
Total deduction	9.00
	$29.50

Thus, 680 lb of copper returns $29.50 = $0.043 per pound.

A perfect partially marmatitic zinc concentrate is assumed to run 60 pct zinc. This would realize:

2000 × 0.60 × 0.8 × $0.0825	$79.20
Total deduction	30.00
	$49.20

Thus, 1200 lb of zinc return $49.20 = $0.041 per pound.

From these data, perfect work therefore assumes the following maximum net returns:

Gold	$32.00 per ounce
Silver	0.675 (95 pct of $0.71) per ounce
Copper	0.043 per pound
Lead	0.039 per pound
Zinc	0.041 per pound

If we assume an ore running as follows:

Au	Ag	Cu	Pb	Zn	Fe
0.20	5.0	0.25	5.0	5.0	10.0

perfect work for this ore would be:

			PER TON
Gold	0.2 × $32.00	=	$6.40
Silver	5.0 × $0.675	=	3.38
Copper	5 lb × $0.043	=	0.21
Lead	100 lb × $0.039	=	3.90
Zinc	100 lb × $0.041	=	4.10
The theoretical maximum return would then be			$17.99

Proof of the accuracy of this method of determining maximum return may be had by resolving the ore into its component "perfect" products:

100 tons of the ore would yield 5.82 tons of 86 pct lead concentrate, assaying 3.44 oz gold, 86 oz silver. This figures:

Gold	3.44 at $32.00	=	$110.00
Silver	86 × $0.675	=	58.00
Lead	2000 × (0.86 – 0.015) × 0.9 × ($0.065 – 0.015)	=	76.00
			$244.00
Total deduction			9.00
			$235.00

Per ton of original ore, the return = $235.00 × 0.0582 = $13.67

100 tons of the ore would yield 0.73 ton of chalcopyrite assaying 34 pct copper and 30 pct iron. This, with its deduction, on the lead schedule, as already shown, realizes $29.50 per ton. Per ton of original ore, the return would be $29.50 × 0.0073 = $0.22.

100 tons of the ore would yield 8.33 tons of 60 pct zinc concentrate. This concentrate, as already shown above, figures to $49.20 per ton. Per ton of original ore, $49.20 × 0.0833 = $4.10.

The total maximum return is then $13.67 plus $0.22 plus $4.10, which totals to $17.99, which is the same figure we arrived at above by the short-cut method.

On the pyrite schedule shown a pure pyrite product, assaying 47 pct iron, would realize only $2.35 per ton for excess iron pay. The deduction of $6.00 for treatment plus freight would then figure to a loss of $3.65 per ton pyrite, or $0.0039 per pound of iron. When a pyrite product has to be made for precious-metal recovery, this negative value for iron may be considered. The $17.99 total would then be decreased by 200 lb × $0.0039, or $0.78, making the maximum return $17.21.

If only a lead concentrate and a zinc concentrate are made on the pattern ore used, the combined return from the two products per ton ore divided by $17.99 would be the metallurgical efficiency

If a lead, zinc, and pyrite concentrate are made, the combined return of the three products per ton ore divided by $17.21 would be the metallurgical efficiency.

It is necessary, of course, to have a 100 pct checkback of all values in order to establish true efficiency. Underruns would lower efficiency below actual and overruns would raise efficiency above actual. It is necessary therefore to calculate to a heads assay by summation. The theoretical maximum return per ton ore is based upon such summated assay. It is extremely important, therefore, to obtain reasonably correct estimates of tonnage, and correct products and tailings assays.

No claim is made that the method of computing metallurgical efficiency does not leave something to be desired. Changing markets influence the figure, of course. But this need not be disturbing, as comparison may be resolved to some one market condition. The difference between efficiency based upon different markets may be too great for comparison, but when such comparison is made at either market, the difference at one market should closely resemble the difference at the other.

Other operators may have developed methods of evaluating mill results as satisfactory as this one. If so, I hope they will give us the benefit of their experience.

DISCUSSION

S. D. Michaelson*—The method of determining the efficacy of a separation described in this paper, gives the metallurgist a most useful guide for keeping his ship on an even keel when the ore waves vary from placid seas to high breakers.

Now that such a method of evaluating metallurgical efficiency has been developed and proven practical, it would seem logical to introduce the next, and more final, factor of *Economic Efficiency*. I choose to define this term as the percentage ratio of net return from milling to the theoretical maximum net return.

The actual operating cost subtracted from the actual mill return represents the net actual amount received for the material treated.

Experienced operators of most plants would ordinarily have enough data on hand to estimate a theoretical milling cost per ton, based on labor, power, supplies and maintenance when operating at maximum tonnage with minimum trouble and under ideal ore conditions. When this theoretical estimated cost is subtracted from the theoretical maximum ore return (calculated by Pallanch's method) the resulting figure is the maximum theoretical net amount the mill operator could hope to realize in processing the ore.

The percentage ratio of these sums, of actual to theoretical, is the *economic efficiency* of the operations.

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* Allis-Chalmers Manufacturing Co., Milwaukee, Wisc.

This *economic efficiency*, if determined on the basis of a fixed market price for metal and similarly fixed base rates for labor, power and supplies, will reflect the effect of changes in milling practice, labor efficiency and equipment condition. When computations made on the basis of actual current rates for labor, power and supplies are compared with similar actual figures for past periods they will show forcefully how energetic our hunt must be to develop man saving controls, more effective metallurgy, and better equipment to maintain profitable operations. Such computations also offer a simple means of determining how much more labor, supplies and power can be put into the mill operation to obtain an increase in metallurgical recovery at a profit.

A Short-cut Method of Metallurgical Accounting

By E. H. Crabtree, Jr.* and Neil S. Parker,* Members AIME

The custom milling plant of the Eagle-Picher Mining and Smelting Co. is at Sahuarita, Arizona, approximately 20 miles south of Tucson. It is connected by a 2-mile railroad spur to the main line of the Southern Pacific Railroad between Tucson and Nogales. The capacity of the plant is 500 tons per day.

The mill is conventional except perhaps in the rather wide variety of ores treated. These vary from straight copper ores, from which only one concentrate is produced, and lead-zinc or copper-zinc ores from which two concentrates are produced, to complex gold-silver-copper-lead-zinc ores, which require the simultaneous production of three or four concentrates.

Custom ores are received in lots varying in size from a few truckloads to several carloads. Similar ores, after crushing and sampling, may be commingled, or individual lots may be milled individually. The ores are first dumped either into one of two 200-ton railroad bins, or into one of the two 75-ton truck bins. From these bins the lot of ore is individually crushed in a 24 by 16-in. jaw crusher to 3-in. size, and then to ½-in. size in a 3-ft Symons shorthead crusher.

After crushing to ½-in., the ore is conveyed to the automatic sampling plant. This consists of three Vezin samplers in series, each successively cutting out 10 pct, 10 pct and 5 pct of its feed. After No. 1 sampler, the material passes through a mixing barrel and is then crushed to ¼-in. in a set of rolls before passing to No. 2 sampler. Between No. 2 and No. 3 samplers, the ore is again mixed in a mixing barrel. The final sample, consisting of one pound per ton of ore, is taken to the laboratory, where it is further crushed to ⅛-in. in a coffeemill, and then riffled down to the size required for grinding in a pulverizer. The capacity of the crushing and sampling plant is up to 100 tons per hour, depending upon the character of the ore.

After sampling, the ore is stored in five ore-storage bins, consisting of four 200-ton bins and one 175-ton bin. Each of these bins is discharged by means of belt feeders and conveyor to the ball mills, so that ores may be composited for milling in any required ratio, or milled separately by themselves.

Fine grinding is done in one 6 by 4-ft Allis-Chalmers ball mill and one 8-ft by 36-in. Hardinge mill. These two mills both discharge to one 54-in. Akins Simplex Highweir classifier. The classifier sands can be returned to either or both mills, so that flexibility in grinding for different ores is easily obtained.

Flotation equipment consists of four 66-in. Fagergren machines for copper-lead roughing and six similar machines for zinc roughing. Lead-copper bulk concentrate is cleaned and recleaned in eight No. 18-S Denver machines, and zinc concentrate is cleaned in three 66-in. Fagergrens. Copper-lead differential separation

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takes place in six 18-S Denvers. Pyrite recovery, when required, is obtained from the zinc rougher tailings by treatment in 18-S Denver machines. Conditioning equipment is provided between the flotation circuits.

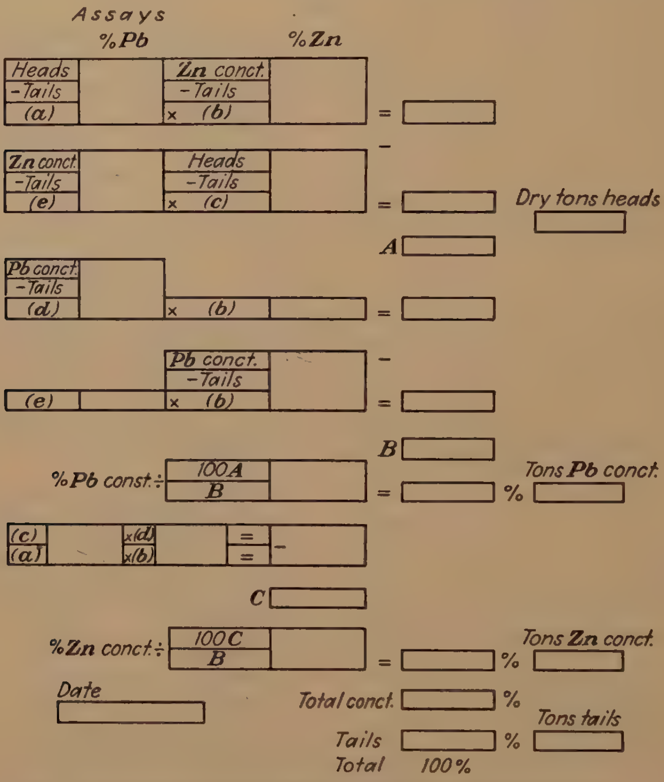

FIG 1—THREE-PRODUCT FORMULA, RATIOS OF CONCENTRATION.

All concentrates are thickened and filtered in American filters, from which they are conveyed to storage. They are then conveyed directly from storage into railroad box cars.

ACCOUNTING

Metallurgical accounting when only one or two concentrates are produced from one ore is relatively simple. It becomes increasingly involved when two or three concentrates are made simultaneously from commingled ores. At Sahuarita, the accounting procedure has been simplified by use of slotted sheets (Figs 1 and 2), which solve the three-product or four-product algebraic formulas as given in Taggart's Handbook of Ore Dressing.

For example, when it is required to know the tonnage of lead and of zinc concentrates produced from a given tonnage of ore of a given assay, the three-product formula sheet is used. In the appropriate slots are inserted the lead and zinc assays of the heads, lead concentrates, zinc concentrates, and tails. Simple cross multiplication, subtraction, and division, as indicated, give the final results in tons of concentrates produced from the individual lot of ore.

When lead, zinc, and copper concentrates are simultaneously produced, the four-product formula sheet is similarly used. The figures representing tonnage of concentrates produced from the individual lots are cumulated throughout the month to determine total monthly production. At the end of each month the concentrate bins are cleaned and a check is obtained on actual mill recoveries. Discrepancies, which are always small, are then allocated back to the individual lots of ore.

These sheets can be used by any competent office accountant without metallurgical knowledge, and it is estimated that many hours of work are saved each month by their use. Comparison of the use of these slotted sheets with the solutions of the formulas from which they are constructed indicates the simplification obtained.

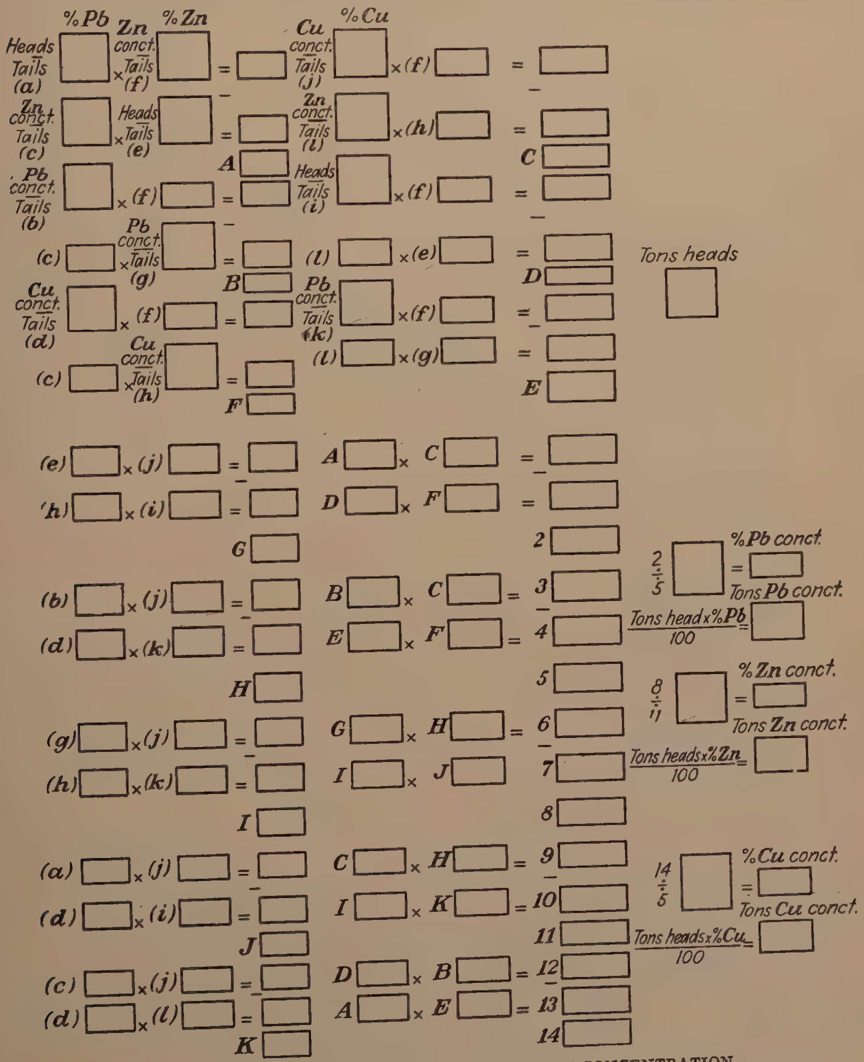

FIG 2—FOUR-PRODUCT FORMULA, RATIOS OF CONCENTRATION.

Crushing Practice at the Braden Copper Company

By E. R. Johnson*

(New York Meeting, March 1947)

The copper concentrator of the Braden Copper Co. is at Sewell, Chile, on the western flank of the main Cordillera of the Andes, at an air distance of approximately 50 miles southeast of Santiago, the capital of Chile. The region is mountainous and space is at a premium because of the rough terrain and the danger of snowslides.

The ore, a porphyry copper, is mined by a block-caving method in the El Teniente mine. It is hauled to the crushing plant, a distance of 2.6 miles, through a tunnel because of the severe climatic conditions. The usual large, coarsely set, primary reduction machines found in many ore-treating plants are not required because of the mining method employed and the limiting 12-in. spacing of the grizzlies at the head of each ore pass in the mine.

It is the purpose of this paper to describe the numerous improvements made in the crushing section of this large flotation concentrator during the past 15 years to increase the efficiency of the plant and to raise the daily output from around 20,000 to 30,000 tons. The reader should keep in mind that all changes were made within the existing building because of the limited space and the high cost of the heavy mill construction required to withstand the snow loads. The present plant, as well as the former, operates on a 24-hr, 7 days a week basis.

FORMER PLANT

The crushing plant commenced operations at its present site in 1921, the main equipment being No. 7½ gyratories, followed by 48-in. horizontal Symons disk crushers. Fines were removed by 1-in. stationary grizzlies ahead of each unit. In 1924 a third stage of crushing was provided by the addition of 55½-in. rolls in open circuit. During 1925, the Symons disks were replaced by 73½-in. rolls. All the rolls were changed to closed circuit, with elevators and screens to give a final product of 9 pct plus 3-mesh.

The flowsheet of a typical crushing section of this period is shown in Fig 1. The ore was withdrawn from a common bin serving the five crushing sections in the plant by means of any number up to six 36-in. wide-apron feeders, which discharged onto a common conveyor belt. This conveyor was equipped with a magnetic head pulley and the crusher attendant was stationed near the discharge end to pick out pieces of mine timber that might be in the ore.

The ore was discharged over a 1¾-in. spaced bar grizzly and the oversize of this grizzly was crushed by a No. 7½ McCully gyratory crusher set at 1¾-in. opening on the closed side. The grizzly undersize joined the gyratory discharge and was conveyed to a set of 73½ by 20-in. Garfield rolls set at 1 in. These rolls operated in closed circuit with a bucket elevator and two vibrating screens placed in parallel.

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* Plains, Montana.

The aperture of these screens was 1¼ in. The screen oversize was returned to the rolls and the undersize was conveyed to four vibrating screens operating in parallel and equipped with 0.338-in. square-opening screen cloth. The oversize of each pair of screens was fed by gravity to a set of 55½ by 20-in. Garfield rolls. The discharge of each set of rolls was returned to the vibrating screens above by means of individual bucket elevators. These finishing rolls were set at 3⁄16 in. between shells.

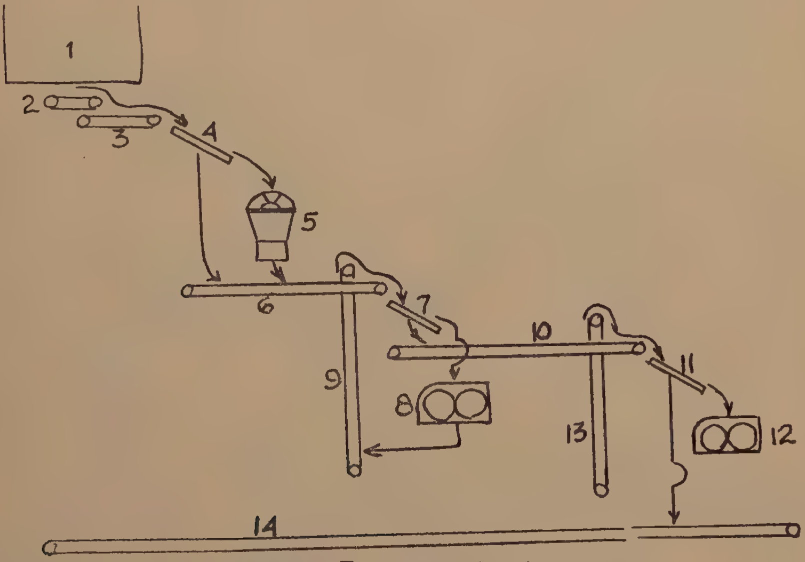

Fig 1—Flowsheet in 1924.

1. Coarse ore bin, 12,500 tons capacity.
2. 36-in. wide apron feeders, six per primary crusher.
3. 48-in. wide wood picking conveyor equipped with magnetic head pulley, one per primary crusher.
4. Tapered bar grizzly, 1¾-in. opening, one per primary crusher.
5. One No. 7½-in. McCully gyratory crusher.
6. One 36-in. wide conveyor.
7. Two 4 by 5-ft Hummer vibrating screens in parallel.
8. One set 73½- by 20-in. Garfield type rolls.
9. One 30-in. bucket elevator.
10. One 36-in. wide conveyor.
11. Four 4 by 5-ft Hummer vibrating screens in parallel.
12. Two sets 55½- by 20-in. Garfield type rolls in parallel.
13. Two 30-in. bucket elevators in parallel.
14. 48-in. belt conveyor collecting finished products from all sections.

Each section of this arrangement was capable of producing about 4000 tons per 24 hr of 9 pct plus 3-mesh material. Delays were numerous and greater capacity was needed in the crushing plant to keep up with the general plant expansion.

Present Plant

The replacement of the No. 7½ McCully crushers by 7-ft. Symons standard cone crushers initiated in 1929 eliminated the intermediate stage of crushing and by simple alterations of the chutes and conveyors, the intermediate stage equipment was used for making a finished product. The flow of ore was as follows:

Ore was withdrawn from the bin as previously and discharged over a tapered bar grizzly having a ¾-in. top opening and a 1¼ in. bottom opening. The oversize fell by gravity into the Symons standard cone crusher and the undersize joined the cone crusher discharge. The primary crusher discharge was split by a mechanical splitter between the set of 73½ by 20-in. rolls and the two sets of 55½ by 20-in. rolls. All the rolls were then arranged to operate in parallel, closed circuits with individual sets of screens and bucket elevators. Although this change increased the section rate by approximately 2300 tons per day, making each section of this arrangement capable of crushing 6300 tons per 24 hr, the full value of the section was never obtained because of insufficient screening capacity in the roll circuits.

At this time a study was made of the screening of the various crusher feeds and products and this led to the installation of two screens in tandem to screen a finished product out of the primary grizzly undersize, as shown in Figs 2 and 3. These screens removed natural fines from the ore amounting to approximately 15 pct of the total plant feed and increased the capacity of each section by an additional 700 tons per 24 hours.

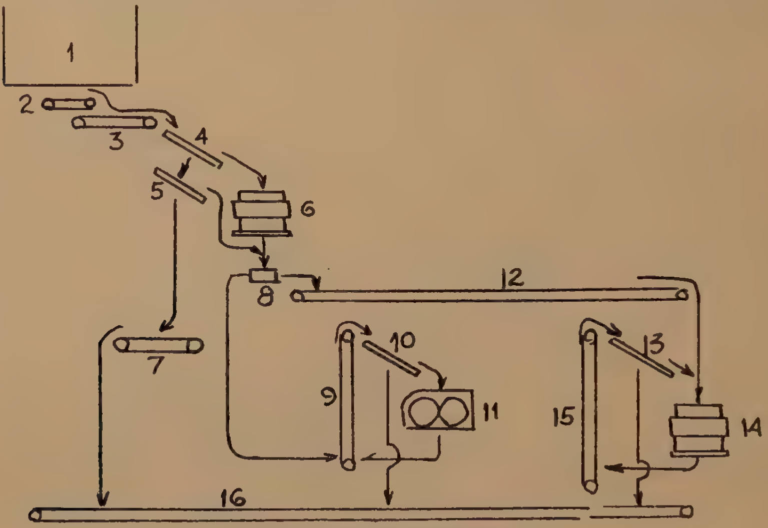

FIG 2—SECOND FLOWSHEET.

1. Coarse ore bin, 12,500 tons capacity.
2. 36-in. wide apron feeders, six per primary crusher.
3. 48-in. wide wood picking conveyor equipped with magnetic head pulley, one per primary crusher.
4. Tapered bar grizzly, ¾-in. top opening, 1¼-in. bottom opening, one per primary crusher.
5. Two 3-ft 10-in. by 5-ft Utah type electromagnetic vibrating screens placed in tandem, one set per primary crusher.
6. One 7-ft Symons standard cone coarse bowl crusher.
7. 36-in. wide collecting conveyor.
8. Mechanical splitter.
9. One 30-in. bucket elevator.
10. Four 3-ft 10-in. by 5-ft Utah type electromagnetic vibrating screens placed in two sets of two each in tandem.
11. One set of 73½ by 20-in. Garfield type rolls.
12. One 36-in. wide belt conveyor.
13. Eight 3-ft 10-in. by 5-ft Utah electromagnetic vibrating screens placed in four sets of two screens each in tandem.
14. One 7-ft. Symons short-head cone crusher.
15. Two 30-in. bucket elevators in parallel.
16. 48-in. belt conveyor collecting finished products from all sections.

Tests were also run using additional screens having the same aperture in the finishing roll circuits. The conclusion arrived at from these tests was that although the section tonnage increased, it increased at the expense of a much larger percentage of material in the coarser sizes. It was decided to obtain the desired increase in plant capacity by other means.

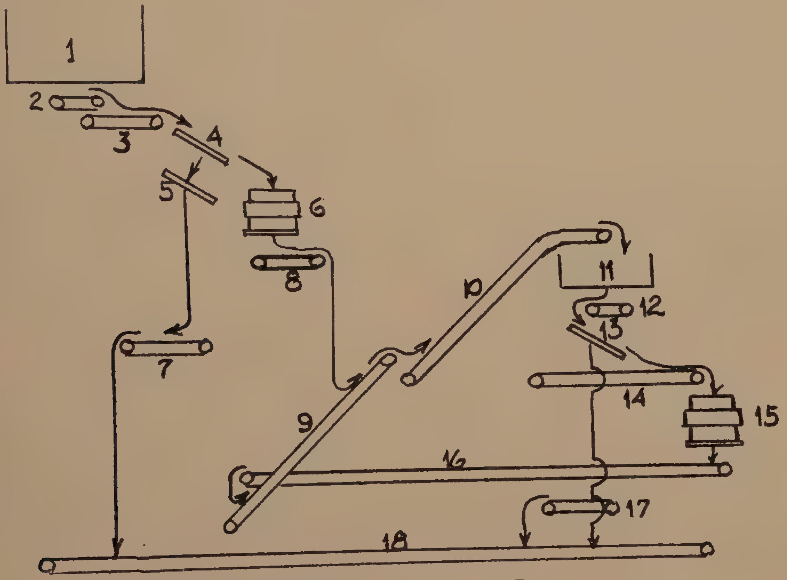

FIG 3—THIRD FLOWSHEET.

1. Coarse ore bin, 12,500 tons capacity.
2. 36-in. wide apron feeders, six per primary crusher.
3. 48-in. wide wood picking conveyor equipped with a magnetic head pulley, one per primary crusher.
4. Tapered bar grizzly, ¾-in. top opening, 1¼-in. bottom opening, one per primary crusher.
5. Two 3-ft 10-in by 5-ft Utah type electromagnetic vibrating screens placed in tandem, one set per primary crusher.
6. Three 7-ft Symons standard cone coarse bowl crushers, two in operation and one spare.
7. 36-in. wide collecting conveyor.
8. 48-in. wide collecting conveyor.
9. 60-in. wide elevating conveyor.
10. 60-in. wide conveyor equipped with traveling tripper car.
11. Surge bin, 1600 tons capacity, serving four secondary crushers.
12. 36-in. wide belt feeders, eight per secondary crusher.
13. 3 by 6-ft Utah type electromagnetic vibrating screens, seven in operation and one spare per secondary crusher.
14. 36-in. wide conveyor feeding each short-head crusher.
15. Four 7-ft Symons short-head cone crushers.
16. 60-in. conveyor collecting discharge from short-head crusher.
17. Two 36-in. conveyors collecting undersize of surge-bin screens.
18. 48-in. belt conveyor collecting finished products from all sections.

A 7-ft Symons short-head cone crusher was purchased in 1937 and installed in one of the sections to operate in parallel with two sets of 55½ by 20 in. rolls, to give a much needed increase in plant capacity. The short-head crusher was arranged to operate in closed circuit by means of two 30-in. bucket elevators and a sufficient number of vibrating screens. It received its share of the combined primary screen oversize and standard cone crusher discharge by the means of a mechanical divider.

Comparisons were made at the time between the rolls and the short head, which resulted in favor of the latter. Now, after nearly nine years of operation of several more short-head crushers, the results are still very much in their favor. All the short-head crushers use "short" type mantles and liners instead of the "long" type commonly used in open-circuit work. It is not practical to quote actual operating costs, which have risen considerably owing to the high cost of material and increased labor charges during the war years, but an idea of the economy effected by the short head can be obtained by comparing the steel consumptions of the two types of crushers. The rolls consumed 0.066 lb per ton crushed as compared to 0.034 lb per ton for the short head. Total power charged to the crushing operation has increased from 2.201 to 2.251 kw per ton but 0.127 kw per ton are now being used to operate the dust-collecting systems.

Distribution of power is shown in Table 1.

TABLE 1—*Distribution of Power*

OPERATION	KW-HR PER TON CRUSHED
Primary crushing only	0.557
Primary crushing auxiliary equipment (feeders, conveyors, etc.)	0.071
Secondary crushing only	1.048
Secondary crushing auxiliary equipment (feeders, screens, conveyors, etc.)	0.441
Dust collectors	0.127
Conveyors carrying crushing-plant product to wet mill	0.007[a]
Total	2.251

[a] This figure is low because the decline-conveyor motors work as generators and throw power back into the system.

Several additional 7-ft units were installed in 1938; each short head replacing two sets of 55½-in. rolls. The flow of the ore through one of these sections is shown in Fig 2. Ore is withdrawn from the bin and crushed by the existing equipment of the primary crushing stage. The discharge of the primary crusher is split mechanically between the short-head circuit and the 73½-in. roll circuit. In the roll circuit, the new feed enters the elevator boot along with the roll discharge and is elevated to the four vibrating screens above the roll. In this manner some of the fines produced in the primary crushing stage are removed before entering the roll. The short head receives its feed directly from the primary crushing stage, no attempt being made to remove the fines produced by the primary crusher. The ore, after being crushed in the short head, is elevated by two bucket elevators to four sets of two screens each in tandem. The oversize of these screens is returned by gravity to the short head and the undersize falls to the conveyor collecting the finished products of all the sections.

In the present plant there are two sections having this arrangement and each section is capable of handling 7000 tons per 24 hr running time. The section is flexible and easy to operate, but there are many delays such as those caused by the patching of screens, cleaning elevator buckets, chutes, and grizzlies that greatly reduce the tonnage output.

In 1940 a plan was approved for installing a surge bin to separate the primary and secondary crushing stages, thereby eliminating many of the delays that lower the output of the existing sections. For several reasons the erection of this bin was delayed until near the end of 1943, when the first part of the bin was placed in operation. As shown in Fig 3, the standard cone Symons receive feed from the oversize of the primary grizzlies. The primary screen oversize joins the primary crusher discharge, which is collected by a common conveyor belt serving the three crushers. Only two of the three primary crushers are required and the third is held as a spare. The primary crusher discharge and primary screen oversize join the secondary or short-head crusher discharge and are elevated by belt conveyors to the top of the surge bin where the ore is delivered as required by a traveling tripper car, which automatically opens and closes the dust covers on top of the bin. Each of the four short-head crushers of the secondary stage of crushing receives feed from eight belt feeders located under the bin. Normally only seven of the eight feeders are used, allowing the remaining feeder and screen to be repaired. Each of the feeders delivers feed to an electromagnetic vibrating screen equipped with 0.328-in. square opening screen cloth. The oversize of these screens is conveyed to a short-head crusher and the undersize is collected by a belt conveyor as a finished product. The short-head crusher discharge is returned to the surge bin by means of 60-in. wide belt conveyors. Another short-head crusher, which has not yet been connected into the surge-bin circuit, receives feed from the primary crushers feeding the surge bin. This short head is operated in closed circuit with bucket elevators and screens as in Fig 2.

The surge-bin section is compact, clean, and easy to operate because of its flexibility and the convenient location of the controls. The principal advantages gained by this installation are increased tonnages due to a 10 pct increase in running time, and the full loading of the secondary crushers.

Typical screen analyses of the various feeds and products of the installation are given in Table 2.

CONVEYORS

Numerous belt conveyors are used in the crushing plant proper and for the transfer of the crusher product to the wet-grinding sections of the concentrator. These belts vary in width from 36 in. wide traveling at relatively slow speeds to belts 60 in. wide traveling at 615 ft per minute. Most of the belts have vulcanized splices, and minor repairs are made with a small, portable vulcanizing unit. Much attention has been paid to maintain the conveyors in good condition. This has paid dividends as one of the 48-in. conveyor belts, 280 ft between centers, and operating at a minus 17° angle has transported more than 60,000,000 tons of the crushing-plant product. This belt will be removed from this service in the near future and used on other conveyors that are not so vital to the plant output.

TABLE 2—*Typical Screen Analyses of Feeds and Products of Surge-bin Installation*

Size	Original Feed	Primary Crusher Discharge	Secondary Crusher Discharge	Combined Feed to Surge Bin	Secondary Crusher Feed
Inch					
On 5.94	6.6				
4.20	5.2				
2.10	14.6				
1.485		3.0		0.4	2.0
1.050	20.9	12.1		3.3	3.8
0.742		25.1	4.7	8.8	7.5
0.525	19.2	19.3	18.8	17.7	24.0
0.371					
Mesh					
On 3	9.6	25.1	44.9	40.7	52.9
4	4.0	4.7	12.1	11.2	6.7
6	2.8	2.3	5.8	5.2	1.5
Through 6	17.1	8.4	13.7	12.7	1.6

Size	Primary Screen Undersize	Surge-bin Screen Undersize	Final Product
Mesh			
On 3	4.3	2.8	3.1
4	11.5	20.4	18.6
6	13.0	19.1	17.8
8	9.3	12.3	11.7
10	8.2	8.2	8.2
14	7.4	6.6	6.8
20	5.8	4.5	4.7
28	4.5	3.2	3.5
35	3.9	2.5	2.8
48	3.0	2.1	2.3
65	2.3	1.7	1.8
100	2.5	1.5	1.7
Through 100	24.3	15.1	17.0

ELECTRICAL SYSTEM

An elaborate electrical interlock system, which requires surprisingly little maintenance, is used whereby "choke-ups" are practically eliminated. The system is so arranged that if any conveyor stops all conveyors and feeders preceding it automatically stop also, but the elevators and crushers still operate. The crushers are also protected by electrically operated thermometers, and oil-pressure switches connected to klaxons to notify the operator in case of trouble.

AUXILIARY EQUIPMENT

The plant is well equipped with overhead traveling cranes that serve all crushing equipment and also the repair bays at the end of each crusher floor. A 15-ton incline hoist serves to transfer material from one floor to another. Among the auxiliary equipment in use in the plant there is an electrical induction heater for heating the 73½-in. roll shells before they are shrunk on the cores, and there are vulcanizers for making endless transmission, conveyor, and feeder belts.

In 1937 the first two large dust-collecting units were installed to remove dust from the entire plant. Much difficulty was experienced with the first units in keeping the collecting pipes clean. In 1943, at the same time the surge-bin section of the plant was erected, provisions were made for the new, large dust-collecting systems designed by the local engineering department. These systems have successfully lowered the dust count well below the permissible limit. The installation has four fans totaling 143,000 cfm at 7.8 in. water gauge. Each is connected to a spray tank that serves one group of ducts. The spray tanks discharge to the ball-mill circuit. In addition to the dust-collecting system, three large semiportable cleaners and one stationary vacuum cleaner are used to remove dust accumulation.

ACKNOWLEDGMENT

The writer wishes to express his gratitude to the officials of the Braden Copper Co. for permission to publish this paper. He especially wishes to thank Mr. L. T. Higgins, Concentrator Superintendent, for his helpful criticisms in preparing it for publication.

Rod Milling—Plant and Laboratory Data

By J. F. Myers,* S. D. Michaelson† and F. C. Bond,‡ Members AIME

(New York Meeting, March 1947)

This work was undertaken with the object of collecting plant data on rod milling, making laboratory tests on representative samples of the various ores, and arriving at a basis for comparing the relative efficiencies of different mills operating on widely different materials and under varying conditions. These efficiencies consider only power consumption and do not include the other elements of operating costs. The over-all economic efficiency may be quite different from the consumption efficiencies with which this paper is concerned. It is hoped that these collected data will be of value in indicating the most favorable conditions of rod-mill operation.

Operating-plant data on 14 different installations from 11 plants throughout the United States and Canada were collected, and laboratory tests were made on samples of these ores. The plant rod mills were all operated wet, and either in open circuit or with circulating loads so small as to be negligible.

Data were obtained on one roll-crushing operation at the Utah Copper Co., for comparison. The rolls were operated wet in closed circuit with 10-mesh vibrating screens, and with a high circulating load averaging about 600 pct. This particular roll operation can be considered as a model of efficiency.

The collected plant data are necessarily incomplete in some instances.

Each ore was assigned a code number, and the name and location corresponding to the code number are listed in Table 1 when permitted.

Table 1—*Plant Code Numbers*

Code No.	Company	Plant Location
A-1	Tennessee Copper Co. London Mines	Copper Hill, Tenn.
A-2	Tennessee Copper Co. London Mines	Copper Hill, Tenn.
B-1	Tennessee Copper Co. Isabella Mine	Copper Hill, Tenn.
C-1		
C-2		
D-1	Waite-Amulet Mines Inc.	Noranda, Quebec
F-1	National Lead Co.	Tahawus, N.Y.
G-1	T.V.A. Fontana Dam	Knoxville, Tenn.
I-39	Reserve Mining Co.	Babbitt, Minn.
I-45	Reserve Mining Co.	Babbitt, Minn.
J-1	St. Joseph Lead Co. Balmat	Emeryville, N.Y.
K-1	Jones & Laughlin Co. Benson Mine	Star Lake, N.Y.
L-1	Kerr-Addison Company	Virginiatown, Ont.
U	Utah Copper Co.	Arthur, Utah

The power data in every instance refer to motor input. Net power was reported by the operator for samples I-39 and I-45, and this was increased by 20 pct to compensate for motor losses and power factor.

The plant rod-mill data are listed in Tables 2 and 3. In Table 2 the speeds are calculated on the basis of partly worn rod-mill liners 2 in. thick. Where *D*

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TABLE 2—*Plant Mill Operation*

Plant Code No.	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
Make	A-C	A-C	A-C	A-C	A-C	A-C	Marcy	A-C	A-C	A-C	A-C	A-C	A-C	Rolls	
Size, in.	6 × 12	6 × 12	6 × 9	9 × 12	9 × 12	9 × 11½	6½ × 12	9 × 12	3 × 6	3 × 6	6½ × 12	9½ × 12	8 × 12	44 × 16	
Speed, rpm	23.9	23.9	24.4	14.0	14.0	16.0	22.0	15.5	35.0	35.0	17.0	14.6	17.0		
Optimum speed, rpm	21.5	21.5	21.5	15.6	15.6	15.6	20.2	15.6	31.9	31.9	20.2	14.8	17.4		
Percentage of critical speed	74.5	74.5	76.4	53.3	53.3	61.2	71.3	59.4	74.5	74.5	54.3	57.2	60.7		64.9
Percentage of optimum speed	111.1	111.1	113.5	89.8	89.8	102.6	108.9	99.4	110.6	110.6	84.2	98.7	97.7		101.6
Peripheral speed, ft per min.	426	426	435	382	382	436	427	423	293	293	330	420	409		390.9
Peripheral speed, average pct	109.0	109.0	111.3	97.7	97.7	111.6	109.2	108.3	74.9	74.9	84.4	107.5	104.8		100.0
Tons per 24 hr.	1,250	2,000	1,050	2,400	3,300	1,250	995	3,276	55.6	115.6	1,004	3,432	1,310	2,597	
Tons per hour	52.1	83.4	43.8	100.0	137.5	52.1	41.5	136.4	2.3	4.8	41.9	143.0	54.6	108.2	
Horsepower input	220	220	172	345	345	300	240	305	23.4	23.4	182	418	236	376	
Kw-hr per ton	3.15	1.97	2.93	2.57	1.87	4.30	4.32	1.67	7.54	3.62	3.24	2.18	3.23	2.59	3.23
Discharge dia., in.	22	22	18	22	22	32	48	39	8	8	24	24	26		
Rod size, in.	3	3	2½	3	3	3	2½		3½	3½	2½	3½	3½		
Dilution, pct solids	77	77	78	75	75	83	65						81		
Liners	2" Lor-aine	2" Lor-aine	2" Lor-aine	Modified wave	Modified wave				Flat	Flat	Wave		Wave		
Backing material	Zn	Zn	Zn	¾" wood	¾" wood	¾' wood	⅛" rubber				⅜" oak		¼" rubber		
Liner wear, lb per ton	0.071	0.053	0.077	0.015	0.011	0.0366	0.125				0.07		0.08		
Rods, pct carbon				1	1	.85	1						1		
Rod wear, lb per ton	0.61	0.46	0.77	0.473	0.334	0.94	1.35				1.077		0.67		
Rods, pct mill volume	45	45	45	38	38	35					60		low		

TABLE 3—*Plant Data, Size, and Surface Area*

Plant Code No.	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
FEED															
Inch:															
1.500	100.0	100.0				100.0		100.0				100.0		100.0	
Percentage passing: 1.052	98.8	98.8		100.0	100.0	98.6		98.9	100.0	100.0		95.0	100.0	92.2	
0.742	92.1	92.1	100.0	95.1	93.5	90.9	100.0	96.5	96.4	96.2		87.9	86.6	68.0	
0.525	85.6	85.6	85.0	86.2	80.1	70.7	83.3	88.0	81.4	78.6	100.0	71.4	64.6	47.1	
0.371	79.0	79.0	59.0	75.8	67.4	53.2	62.0	76.0	64.0	64.5	96.4	59.5	52.3	34.1	
Mesh:															
3	72.1	72.1	46.3	66.6	57.7	41.6	45.3	65.3	51.7	53.2	84.8	49.9	42.9	27.1	
4	65.2	65.2	38.2	57.8	49.6	33.6	37.8	56.0	42.3	42.6	70.6	44.9	35.3	22.3	
6	60.2	60.2	33.9	51.1	43.2	29.0	32.1	48.1	34.5	34.9	61.0	40.0	29.7	18.7	
8	55.6	55.6	29.6	45.1	38.3	24.6	27.8	41.0	28.0	28.2	54.1	36.6	25.1	16.4	
10	50.8	50.8	27.0	39.0	32.7	21.3	23.8	35.9	23.0	23.0	47.5	33.0	20.7	14.6	
14	46.3	46.3	23.3	34.7	29.0	18.6	20.7	29.0	18.5	18.7	40.9	30.5	17.3	13.1	
20	40.0	40.0	21.6	30.2	26.1	16.3	17.4	25.4	15.0	15.1	35.0	27.1	15.0	11.9	
28	34.7	34.7	18.7	27.6	23.2	14.4	15.8	21.0	12.2	12.5	29.6	23.3	13.1	10.8	
35	29.5	29.5	16.0	24.1	20.2	12.9	13.5	17.8	9.9	10.5	24.3	18.9	11.6	9.8	
48	23.5	23.5	13.8	21.8	18.3	11.5	11.1	15.0	8.0	8.1	20.0	15.1	10.3	8.9	
65	18.7	18.7	11.5	19.9	16.7	10.3	9.0	12.5	6.7	6.6	15.7	11.3	9.2	7.9	
100	14.0	14.0	9.3	18.0	15.1	8.7	7.1	10.7	5.6	5.4	11.6	7.8	8.2	6.8	
150	10.8	10.8	7.8	17.0	14.3	7.2	5.3	9.1	4.5	4.4	8.8	5.7	7.4	5.7	
200	8.6	8.6	6.5	15.5	13.1	6.0	4.0	6.8	3.7	3.6	6.4	3.7	6.7	4.9	
Slope *m*	0.727	0.727	0.519	0.303	0.318	0.518	0.757	0.493	0.570	0.602	0.859	1.06	0.434	0.500	0.599
Intercept *K*	2,200	2,200	14,600	41,500	50,000	17,000	5,000	13,500	23,500	18,500	1,800	1,600	70,000	30,000	
Surface: Bond	5.80	5.80	5.29	13.40	11.50	4.90	2.87	6.40	2.95	2.78	4.12	2.45	4.40	4.10	
Coghill	91.8	91.8	58.5	105.0	88.7	51.1	45.6	66.7	38.9	38.8	78.6	54.9	50.0	39.4	
Tyler	77.1	77.1	49.8	90.9	77.1	43.4	38.0	56.1	32.6	32.4	65.2	45.1	42.9	33.5	
PRODUCT															
Mesh:															
0.371													100.0		
Percentage passing: 3								100.0		100.0			99.7		
4					100.0	100.0		99.3		99.0			98.7		
6				100.0	97.8	99.1		96.7		98.3		100.0	96.7	100.0	
8				99.5	94.3	97.0	100.0	92.0		95.5		99.0	88.0	98.8	
10	100.0	100.0	100.0	97.4	84.6	92.3	99.8	84.9	100.0	89.0	100.0	97.1	77.0	90.4	
14	99.9	98.9	98.9	92.3	74.5	85.1	99.1	73.0	99.0	73.7	99.9	92.6	65.5	81.5	
20	99.4	96.9	96.2	84.6	65.1	75.1	95.7	63.3	98.0	60.0	99.4	85.5	56.9	73.6	
28	97.3	91.7	92.2	73.3	55.7	66.5	92.1	51.5	95.5	49.0	97.1	74.6	48.9	64.9	
35	90.1	80.0	81.2	60.9	45.7	57.3	82.5	43.7	83.0	40.0	88.7	61.0	42.7	57.4	
48	75.5	64.5	69.2	53.1	40.1	51.2	69.0	35.0	67.6	33.2	76.6	49.4	37.0	51.6	
65	61.8	51.2	56.2	47.2	35.9	43.7	56.0	29.0	55.9	26.3	60.7	37.7	32.3	45.5	
100	46.2	38.3	44.1	41.4	32.1	36.8	44.2	23.7	45.4	21.5	44.0	27.3	28.3	39.7	
150	37.1	30.0	36.2	38.5	30.1	31.2	33.3	21.0	38.0	17.9	32.5	20.4	25.3	33.7	
200	28.8	22.9	29.6	34.1	27.2	24.5	24.5	15.7	30.4	14.1	23.3	15.3	22.3	28.9	
Slope *m*	0.695	0.777	0.587	0.339	0.340	0.503	0.541	0.552	0.576	0.593	0.937	1.07	0.374	0.460	0.596
Intercept *K*	410	490	590	1700	3500	1150	355	1900	550	1950	345	655	4000	1150	
Surface: Bond	20.00	14.20	22.00	31.00	24.30	20.50	32.50	13.25	23.60	10.70	12.40	5.31	20.00	23.70	
Coghill	265.5	229.2	254.7	240.1	190.3	205.9	244.3	149.3	259.5	138.8	249.9	175.9	166.9	219.0	
Tyler	223.9	192.4	216.5	206.6	163.9	174.8	205.1	125.8	220.2	116.7	208.8	146.5	143.0	187.5	
New Surface:															
Bond	14.20	6.70	16.71	17.60	12.80	15.60	29.63	6.85	20.65	7.92	8.28	2.86	15.60	19.60	
Coghill	173.7	137.4	196.2	135.1	101.6	154.8	198.7	82.6	220.7	100.0	171.3	121.0	116.9	179.6	
Tyler	146.8	115.3	166.7	115.7	86.8	131.4	167.1	69.7	187.6	84.3	143.6	101.4	100.1	154.0	
Square meters surface per kw-hr, Bond	9,050	8,560	11,190	23,400	23,400	9,828	13,970	14,050	7,260	5,810	6,700	3,680	15,820	25,400	
Joules per sqm: Bond	398	421	322	154	154.0	366	258	256	495	620	5.38	978	228	14.2	386
Mesh:															
Kw-hr per net ton: −14	5.88	3.74	3.87	4.46	4.11	6.46	5.50	3.79	9.36	6.58	5.49	3.51	6.70	3.79	5.23
−48	6.05	4.80	5.29	8.22	8.58	10.82	7.46	8.35	12.65	4.40	5.72	6.35	12.10	6.07	8.34

equals the mill diameter inside the liners in feet, the optimum rod-mill speed is calculated from formula 1:

$$\text{Optimum speed, rpm} = 0.80(57 - 40 \log D) \quad [1]$$

in which it is assumed that the optimum rod-mill speed is 80 pct of the optimum ball-mill speed. The optimum per cent is the percentage ratio of actual speed to calculated optimum speed. This formula was derived to agree with the average recommended speeds for a large number of different diameters in different installations.

The critical speed is calculated from the formula

$$\text{Critical speed, rpm} = 76.63/\sqrt{D} \quad [2]$$

The actual operating speeds conform more closely to the optimum speed formula than to a constant percentage of critical speed. The peripheral speeds in feet per minute inside the liners, and the percentage of the average peripheral speed, are listed.

Calculations of Surface Area

The screen analyses of the plant feed and products are evaluated in Table 3. The surface areas are calculated according to the Bond,[2] Tyler,[3] and Coghill[4] methods.

In the Bond method the weight per cent passing is plotted against the screen openings in microns on log-log paper, and a straight line is drawn corresponding to the plotted line. The intercept K of this straight line in microns at 100 pct passing, and the slope of m of the line, are used to calculate the surface area of all particles larger than 0.7 micron according to the following equation:

$$S = \frac{600m}{K(1 - m)}\left[\left(\frac{K}{L}\right)^{1-m} - 1\right] \quad [3]$$

where S is the surface area in square meters per 100 cc of solid.

Equation 3 has been plotted so that the surface areas can be found directly from K and m without calculation. L equals 0.7.

In the Coghill method the surface area is calculated by multiplying the weight per cent retained on any one sieve in the Tyler standard screen scale and passing the next larger sieve, by the reciprocal of the screen opening in inches. The percentage passing 200-mesh is considered to be all retained on the next finer size of 270-mesh. The total surface area is the sum of these products, representing 100 pct of the sample, in Coghill surface units. This quantity multiplied by tons of ore is usually designated as Coghill surface-tons. The surface units are arbitrary, and express only relative values for comparative surfaces. In calculating the reciprocals a 1.050-in. opening testing sieve is considered to be 1.000 unit; a ¾-in. screen, 1.414 units; a ½-in. screen, 2.00 units; and so on, increasing by $\sqrt{2}$ for each successively smaller screen size.

The Tyler method is similar to the Coghill except that the actual average linear opening of the retaining screen and that of the next larger screen passed is used. The reciprocal value of material passing one inch and retained on ¾ in. is 1.114, and the finer sizes increase in the $\sqrt{2}$ ratio. The reciprocal value of material passing 150-mesh and retained on 200-mesh is 285.714. The percentage passing 200-mesh is assigned the value of 454.545. Units are expressed as in the Coghill method.

The Tyler method gives slightly lower values than the Coghill method. It is obvious that the bulk of the minus 200-mesh surface area is largely ignored in both methods, since it is assumed to be all plus 270-mesh in the Coghill method, and all plus approximately 325-mesh in the Tyler method. The actual work done in grinding particles finer than these sizes is an unavoidably large part of the total work input to the mill, even though the production of these fine particles may not be desired.

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[2] References are at the end of the paper.

Since the surface area per unit volume equals 6/diameter, the Tyler surface area in square meters per 100 cc of solid equals $^{6}/_{254}$ multiplied by Tyler surface tons, since 25.4 mm equals one inch.

The surface area in square meters per 100 cc of solid according to the Coghill method equals Coghill surface tons multiplied by $^{6}/_{254} \times \sqrt[4]{2}$.

The total Coghill surface is slightly more than the Tyler in spite of the larger average particle size assumed for the minus 200-mesh portion.

The new surface area is found by difference, and the square meters of new surface per kilowatt-hour and the joules input required to produce one square meter of new surface are calculated for Bond surfaces. In all three methods the surface areas are calculated for spherical particles.

Relative Grinding Efficiencies

At the bottom of Table 3 are listed the kilowatt-hours required per net ton of minus 14-mesh and per net ton of minus 48-mesh produced in the plant mills. Since the grind in some instances was about minus 20-mesh the power required per net ton of minus 48-mesh probably is more representative than that of the minus 14-mesh.

In Table 4 the laboratory rod-mill grindability tests[5] are tabulated. The screen analysis of the laboratory rod-mill feed and products, and the Bond, Coghill, and Tyler surface areas, are listed as in Table 3. The 14-mesh grindability, or net grams of minus 14-mesh (1190 microns) produced per revolution of the rod mill, is tabulated, together with the gross grams per revolution. The gross or total grams per revolution is necessary to calculate the new surface area produced.

The joules per square meter of surface for each ore was computed on the assumption that one revolution of the laboratory rod mill is equivalent to 83.5 joules of net energy going into the ore sample. It has been previously determined that one revolution of the standard laboratory ball mill represents an energy of 52.3 joules of useful work done on the ore.[2] The value of 83.5 joules per revolution of the rod mill was obtained by averaging and comparing ball-mill and rod-mill grindability results on a number of different ores.

In Table 5 the relative grinding efficiencies of the various ores are calculated on the basis of "product screen size" production, rather than on the basis of surface areas. The product size passed by 80 pct of the plant rod-mill products was selected as the basis of comparison, and this size in microns is listed in the first line of the table. The size in microns passed by 80 pct of the rod-mill feed is listed in the second line; and the reduction ratio at the size passed by 80 pct is listed in the third line. The percentage passing the product size in the plant rod-mill feed is listed in the fourth line The kilowatt-hours per ton used by the plant mill divided by the percentage of product-size material in the mill discharge, minus the percentage of product-size material in the mill feed, equals the kilowatt-hours per net ton of product-size material made. This is listed in the fifth line of Table 5.

In the sixth line the equivalent rod-mill grindabilities for each ore at the product size of the plant grind are listed. This was obtained by plotting the standard grindability at 14-mesh on the log-log screen-analysis plot and extending this line parallel to the plotted percentage passing line of the ground product. It is known from numerous tests that the plotted grindabilities have approximately the same slopes as the plotted percentage passing lines of the corresponding ground products.

TABLE 4—*Laboratory Rod-mill Grindability*

Plant Code No.	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
Specific gravity	4.52	4.52	4.63	2.65	2.65	3.35	4.46	2.65	3.42	3.42	3.46	3.23	2.77	2.70	
FEED															
Inch:															
1.500						100.0									
Percentage passing: 1.052			100.0	100.0	100.0	99.6			100.0	100.0		100.0		100.0	
0.742	100.0	100.0	98.6	98.6	98.6	97.2	100.0	100.0	98.5	98.5		99.0		97.9	
0.525	88.0	88.0	78.0	91.1	91.1	80.6	85.2	97.1	79.6	79.6	100.0	90.3	100.0	85.2	
0.371	68.0	68.0	53.8	79.4	79.4	58.5	47.8	83.9	57.6	57.6	96.4	73.4	65.6	58.4	
Mesh:															
3	55.7	55.7	37.4	65.6	65.6	40.7	28.5	67.9	38.3	38.3	84.8	63.5	47.4	40.7	
4	49.3	49.3	31.8	59.5	59.5	32.8	20.2	59.7	30.6	30.6	70.6	53.5	34.2	30.5	
6	45.7	45.7	26.8	49.6	49.6	26.9	16.4	41.8	23.5	23.5	61.0	47.9	26.0	25.1	
8	42.1	42.1	24.0	42.4	42.4	22.9	13.6	45.0	18.9	18.9	54.1	43.3	20.9	20.9	
10	39.0	39.0	21.7	37.3	37.3	20.1	11.7	37.5	15.3	15.3	47.5	39.8	16.6	17.8	
14	35.7	35.7	20.2	33.5	33.5	17.9	10.2	29.1	12.7	12.7	40.9	35.7	13.0	15.3	
20	31.7	31.7	18.4	29.1	29.1	16.2	8.9	22.9	11.1	11.1	35.0	31.4	10.7	13.4	
28	27.5	27.5	16.6	26.6	26.6	14.7	7.8	18.0	9.4	9.4	29.6	26.7	8.8	11.9	
35	23.3	23.3	14.7	23.6	23.6	13.4	6.7	14.3	7.9	7.9	24.3	21.5	7.2	10.5	
48	19.5	19.5	12.9	21.1	21.1	12.2	5.7	11.7	6.6	6.6	20.0	16.9	5.9	9.2	
65	15.9	15.9	11.0	26.3	26.3	8.7	4.7	9.6	5.3	5.3	15.7	12.3	4.8	7.8	
100	12.8	12.8	9.1	12.0	12.0	7.5	3.8	7.9	4.2	4.2	11.6	8.3	4.0	6.5	
150	10.4	10.4	7.5	9.9	9.9	6.7	2.9	6.5	3.2	3.2	8.8	5.8	3.4	5.2	
200	8.3	8.3	5.5	6.2	6.2	5.0	2.2	5.0	2.5	2.5	6.4	3.8	2.9	3.4	
Slope *m*	0.605	0.605	0.473	0.819	0.819	0.500	0.757	0.615	0.678	0.678	0.840	1.095	0.535	0.572	0.685
Intercept *K*	4,500	4,500	24,000	1,900	1,900	24,000	12,000	10,000	16,500	16,500	1,925	1,450	57,000	18,500	
Surface: Bond	6.30	6.30	6.50	4.55	4.55	4.60	1.50	3.80	1.92	1.92	4.15	2.48	2.30	3.35	
Coghill	79.34	79.34	52.94	75.97	75.97	48.63	25.07	55.32	29.44	29.44	78.54	59.36	30.03	39.04	
Tyler	66.77	66.77	45.56	64.22	64.22	40.05	21.11	46.13	24.46	24.46	65.20	49.15	25.13	32.46	
PRODUCT															
Mesh:															
10	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	
Percentage passing: 14	98.1	98.1	99.3	99.4	99.4	97.2	99.3	96.9	97.3	97.3	99.9	99.4	99.6	99.1	
20	81.1	81.1	81.3	78.4	78.4	79.7	78.3	71.0	76.9	76.9	99.4	79.6	76.4	83.8	
28	66.6	66.6	67.4	64.3	64.3	65.7	61.8	53.8	62.7	62.7	97.1	62.2	60.1	70.4	
35	54.0	54.0	55.7	53.8	53.8	54.6	48.5	41.5	51.5	51.5	88.7	46.5	48.2	59.1	
48	43.7	43.7	46.3	46.2	46.2	46.6	38.2	33.3	42.1	42.1	76.6	34.8	39.9	50.7	
65	34.5	34.5	37.8	39.9	39.9	39.5	29.2	26.9	34.3	34.3	60.7	24.4	33.5	42.8	
100	26.8	26.8	30.5	34.4	34.4	33.0	21.9	22.0	27.1	27.1	44.0	16.2	28.6	35.4	
150	20.9	20.9	24.9	30.0	30.0	26.8	15.6	18.2	21.4	21.4	32.5	11.4	24.9	29.4	
200	15.9	15.9	19.9	25.8	25.8	18.3	10.2	14.3	16.7	16.7	23.2	7.9	22.0	24.2	
Slope *m*	0.605	0.605	0.475	0.425	0.425	0.497	0.512	0.610	0.674	0.674	0.865	1.07	0.427	0.503	0.598
Intercept *K*	960	960	1000	1750	1750	1300	910	1750	1000	1000	800	750	2750	1150	
Surface: Bond	15.70	10.70	24.30	22.50	22.50	19.70	22.50	10.80	12.00	12.00	7.60	4.71	18.50	20.50	
Coghill	170.10	170.10	186.14	203.16	203.16	188.01	144.72	146.12	168.02	168.02	155.71	128.40	179.83	207.67	
Tyler	142.10	142.10	157.24	173.12	173.12	157.47	119.30	122.42	141.20	141.20	129.03	105.20	152.82	175.89	
New surface:															
Bond	9.40	9.40	18.80	17.95	17.95	15.10	21.00	7.00	10.08	10.08	3.45	2.23	16.20	17.15	
Coghill	90.76	90.76	133.20	127.19	127.19	139.38	119.65	90.80	139.58	139.58	77.17	691.04	149.80	168.63	
Tyler	75.33	75.33	111.68	108.90	108.90	117.42	98.19	76.29	115.12	115.12	63.83	56.05	127.69	143.43	
Grindability:															
Net	31.2	31.2	16.8	14.7	14.7	6.8	12.2	19.2	8.9	8.9	22.4	28.5	8.8	11.9	16.9
Gross	48.6	48.6	21.1	22.1	22.1	8.3	13.6	27.1	10.2	10.2	37.9	44.3	10.1	14.0	24.2
Joules per sq m: Bond	82.6	82.6	97.4	55.8	55.7	223.0	130.0	116.5	278.0	278.0	221.0	273.0	141.0	94.0	152.1

TABLE 5—*Plant Size Production Efficiency*

Plant Code No.	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
Product size, microns (80 pct passes)	335	400	400	750	1,500	1,000	380	1,500	380	1,300	320	700	1,700	1,100	840
Feed size, microns (80 pct passes)	10,000	10,000	12,000	10,500	13,000	15,500	12,500	10,000	12,500	13,000	6,000	15,500	17,000	22,000	
Reduction ratio at plant speed	29.9	25.0	30.0	14.0	8.7	15.5	32.9	6.7	32.9	10.0	18.8	22.1	10.0	20.0	19.71
Percentage passing P.S. in feed	25.0	29.0	15.8	29.4	32.0	17.5	13.0	34.0	9.5	21.0	21.0	25.0	22.0	12.9	21.9
Kw-hr per net ton of P.S.	5.73	3.86	4.57	5.08	3.89	6.88	6.45	3.63	10.70	6.14	5.49	3.96	5.57	3.86	5.415
Grindability at P.S.	14.5	17.0	10.1	12.2	16.3	6.2	7.2	23.0	4.2	9.5	7.6	16.3	10.5	11.2	11.8
Relative P.S. efficiency	1.20	1.52	2.16	1.61	1.58	2.34	2.15	1.20	2.22	1.71	2.40	1.55	1.71	3.21	1.83
Relative minus 100-mesh efficiency	0.328	0.396	0.707	0.619	0.618	0.961	0.704	0.405	0.597	0.500	0.446	0.314	0.314	1.07	0.570

The relative grinding efficiency is listed in the seventh line of Table 5. This is equal to 100 multiplied by the reciprocal of the product of the kilowatt-hours per net ton of product size (line 5) and the grindability at product size (line 6). This appears to be the most practicable method of measuring the relative plant grinding efficiency on a screen-size basis of different mills grinding different ores to different sizes.

The eighth line of Table 5 lists the calculated relative efficiencies of producing minus 100-mesh material. It is the reciprocal of the product of kilowatt-hours per net ton of minus 100-mesh by the rod-mill grindability at minus 14-mesh.

The plant grinding efficiencies are compared in Table 6. In the first line the efficiency of the production of new surface area is listed on the basis of Bond surface. These values are obtained by dividing the joules per square meter, from the laboratory grindability test in Table 4, by the joules per square meter required in the plant, as listed in Table 3. The comparative efficiencies on the basis of equivalent Coghill surface are listed in the second line, and those on the basis of equivalent Tyler surfaces in the third line.

The relative product-size efficiencies have been recalculated for direct comparison with the Bond surface area efficiencies. This was done by multiplying each relative product-size efficiency by the ratio of the average of all Bond surface-area efficiencies (Table 6, line 1) to the average of all the relative product-size efficiencies (Table 5, line 7). These are listed in the fourth line of Table 6.

Comparison of the last line with the first line of Table 6 will indicate the difference in the efficiency of grinding on the basis of product screen size and that of grinding on the basis of producing new surface area.

TABLE 6—*Comparative Efficiencies*

	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
Surface area efficiency, pct: Bond	20.7	19.6	30.2	36.2	36.2	60.9	50.4	45.5	56.2	44.8	41.1	27.9	61.8	66.2	42.69
Eq. Coghill	26.1	33.3	49.3	39.0	40.0	65.3	59.6	42.4	43.4	41.2	37.7	38.1	50.0	46.4	43.7
Eq. Tyler	25.6	32.7	49.2	38.9	40.1	65.3	59.7	42.7	43.1	40.7	38.1	36.9	49.7	47.9	43.6
Product-size efficiency	27.9	35.4	50.3	37.5	36.8	54.5	50.1	27.9	51.7	39.8	55.9	36.1	39.8	53.8	42.69

TABLE 7—*Impact Crushing Tests*

	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
Average	6.43	6.43	10.36			13.86	24.30	12.30	18.81	18.81		9.84		12.58	13.73
Maximum	8.80	8.80	16.00			24.30	30.80	21.90	25.6	25.6		15.5		19.8	20.20
Plant kw-hr per ton	3.15	1.97	2.93	2.56	1.87	4.30	4.31	1.66	7.54	3.62	3.23	2.18	3.22	2.59	3.23
100 grindability at 14 mesh	3.21	3.21	6.14	6.80	6.80	14.70	8.20	5.20	11.20	11.20	4.47	3.51	11.40	8.40	7.45

TABLE 8—*Abrasion Tests*

	A-1	A-3	B-1	C-1	C-2	D-1	F-1	G-1	I-39	I-45	J-1	K-1	L-1	U	Average
Abrasion Index, Mg	37.60	37.60	32.65	2.65	2.65	8.55	49.85	18.08	29.68	28.68			14.08	5.10	
Rod wear: lb per net ton −48 mesh	1.19	1.12	1.89	1.51	1.53	2.37	2.34				1.91		2.51		
Lb per net ton −200 mesh	3.02	3.22	3.33	2.55	2.37	5.08	6.59				6.38		4.29		

OTHER TESTS

Impact crushing tests[6] were made on ores that contained plus 2-in. pieces in the sample; they are listed in Table 7. The impact value is the number of foot-pounds per inch of thickness between the impact hammers required to break a minus 3-in. plus 2-in. stone. The average of 10 pieces broken is listed in the first line and the maximum value, or hardest piece, in the second line.

The average of the different ores is 13.73 and the average of the maximum values is 20.20 ft-lb per inch. For comparison, the plant rod-mill power consumption in kilowatt-hours per ton is listed in the third line and 100 ÷ the rod-mill grindability at 14 mesh in the fourth line.

TABLE 9—*Order of Decreasing Product-size Efficiency*

Code	P.S. Eff.	Bond Eff.	Impact Strength	14-mesh Grindability	Lab. Slope *M*	Mill Dia., Ft	Mill Length, Ft	Speed Rpm	Critical Speed, Pct	Optimum Speed, Pct
J-1	55.9	41.1		22.4	0.865	6½	12	17.0	54.3	84.2
D-1	54.5	60.9	13.86	6.8	0.497	9	11½	16.0	61.2	102.6
U	53.8	66.2	12.58	11.9	0.503					
I-39	51.7	56.2	18.81	8.9	0.674	3	6	35.0	74.5	110.6
B-1	50.3	30.2	10.36	16.8	0.475	6	9	24.4	76.4	113.5
F-1	50.1	50.4	24.30	12.2	0.512	6½	12	22.0	71.3	108.9
I-47	39.8	44.8	18.81	8.9	0.674	3	6	35.0	74.5	110.6
L-1	39.8	61.8		8.8	0.427	8	12	17.0	60.7	97.7
C-1	37.5	36.3		14.7	0.425	9	12	14.0	53.3	89.8
C-2	36.8	36.2		14.7	0.425	9	12	14.0	53.3	89.8
K-1	36.1	27.9	9.84	28.5	1.07	9½	12	14.6	57.2	98.7
A-3	35.4	19.6	6.43	31.2	0.605	6	12	23.9	74.5	111.1
A-1	27.9	20.7	6.43	31.2	0.605	6	12	23.9	74.5	111.1
G-1	27.9	45.5	12.30	19.2	0.610	9	12	15.5	59.4	99.4
Ave.	42.69	42.69	13.73	16.9	0.598				64.9	101.6

The results of preliminary abrasion wear tests are listed in Table 8. In those tests 500 grams of minus 14-mesh ore are placed in a cylindrical glass jar 3½ in. inside diameter and 6½ in. long, sand-blasted on the inside to prevent slippage, with three weighed black Bakelite plugs 1¼ in. in diameter and ½ in. long, and rotated on a rolling table at 24 rpm for 24 hr. The average milligrams weight loss of the Bakelite plugs is the abrasion index. Only a few abrasion wear tests have been made as yet, and data for interpretation of these results are lacking.

The reported plant rod wear in pounds per net ton of minus 48-mesh and of minus 200-mesh is listed in Table 8.

Standard ball-mill grindability tests[2] were made on some of these ores as follows: at 48-mesh A-1, 8.33; B-1, 4.72; G-1, 3.69; J-1, 5.58; and at 200-mesh K-1, 0.87.

In Table 9 the various ores tested are listed in the order of decreasing product-size efficiency. Comparison with the various other quantities listed shows that in most cases it is necessary to plot the values before any definite trends become apparent.

The various quantities were plotted against each other, and the following trends were established from the plots:

Product-size Efficiency:

Increases as impact crushing strength increases.

Increases as Bond surface production efficiency increases.

Increases as minus 100-mesh production efficiency increases.

Increases slightly as mill diameter decreases.

Increases slightly as product size in microns decreases.

Is without trend as reduction ratio increases.

Is without trend as percentage of critical speed increases.

Is without trend as percentage of optimum speed increases.

Is without trend as mill volume increases.

Bond surface area production efficiency:

Increases as impact crushing strength increases.

Increases as product-size efficiency increases.

Increases as minus 100-mesh production efficiency increases.

Increases slightly as mill diameter increases.

Increases slightly as peripheral speed decreases.

Increases slightly as product size in microns increases.

Increases slightly as reduction ratio decreases.

Increases slightly as mill volume increases.

Is without trend as percentage critical speed increases.

Is without trend as percentage optimum speed increases.

Minus 100-mesh efficiency:

Is without trend as reduction ratio increases.

Is without trend as product size in microns increases.

Average impact strength:

Increases as minus 14-mesh rod-mill grindability decreases.

Increases as surface energy in joules per square meter increases.

SUMMARY AND CONCLUSIONS

Thirteen wet rod-milling and one roll-crushing operations, in which different ores were ground to different sizes under different conditions, were tested and tabulated. Laboratory rod-mill grindability tests were made on each ore, and the plant grinding efficiencies were calculated. The plant grinding efficiencies were calculated on the basis of electrical energy required: (1) per unit of new surface area produced in grinding, and (2) per net unit of product-size material produced. Product-size material is defined as material smaller than the micron particle size passed by 80 pct of the mill product. The grinding efficiencies listed refer only to power consumption, and not to the total cost of grinding.

The operating data supplied were deficient in several instances, particularly in the weight of the rod charge used in the mills. The variations in the calculated results were large, as might be expected from comparisons between grinding widely different ores to different sizes under different conditions, so that results undoubtedly would have been more conclusive if a larger number of installations had been considered.

However, definite indications were obtained from the plotted test results, which are summarized below:

1. Rod mills are more efficient when grinding harder ores; that is, ores with high impact crushing strengths, low grindabilities, and high surface energies.

2. Installations that are more efficient in grinding to pass a certain mesh size (product size efficiency) are also generally more efficient in producing new surface. However, this relationship is not exact; for instance, the Utah Copper Company's crushing rolls were the most efficient producers of new surface of all the installations tested, but were third in the list in the production of mesh undersize. The efficiency of the production of minus 100-mesh undersize also generally follows that of surface area and product size.

3. The effect of mill diameter on the efficiency is small, and is not definitely shown by the tests. The indications are that an increase in the mill diameter slightly decreases the product-size efficiency

and increases the surface-area efficiency, but these cannot be regarded as conclusive.

4. Grinding to a finer size apparently slightly increases the product-size efficiency and decreases the surface-area production efficiency. However, these trends are also too small to be conclusive.

5. An increase in the reduction ratio on the installations tested apparently does not affect the product-size efficiency and slightly decreases the surface-area production efficiency.

6. The mill speed of the installations tested apparently does not affect either the product-size or surface-area production efficiency.

7. An increase in the mill volume also apparently does not affect the product-size efficiency and the surface-area production efficiency.

8. All these comparisons except No. 1 could be made more exactly on different installations grinding the same ore. However, these are not available in most instances.

9. Variations in tonnage capacity at product size with mill diameter, volume, speed, and other factors could be of considerable interest. However, these comparisons were not made because the size of the rod charge was not reported for seven of the mills listed, and the six reported values varied from 35 to 45 pct of the mill volume.

Because of the large number of variables concerned, some of these indications may be obscured by other unconsidered factors or combinations of factors.

Surface-area production efficiency is less affected by the mill-operating variables than the size production efficiency. This would be expected from Rittinger's law. However, the size production efficiency is probably of more immediate concern to mill operators.

The authors are indebted to the various mill operators for contributions of operating data and test samples; also to the several recognized authorities on grinding who have offered suggestions and constructive criticism during the progress of this work.

It is intended to extend this investigation as other data become available, and possibly to investigate the field of ball milling in a similar manner.

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Supplement

J. F. Myers, S. D. Michaelson and F. C. Bond—Some of the operators to whom we sent advance copies of the paper requested that we include a sample calculation for them to follow in figuring efficiencies at their own installations. Rather than delay publication of the original paper we decided to submit such an example as part of the written discussion.

Determination of rod mill comparative efficiency requires data from a standard grindability test on each ore being investigated. The procedure for making this laboratory test is described by Fred C. Bond and Walter L. Maxson.[5] If the operator does not have the necessary equipment for running such a test, a 250 lb sample cut from plant rod-mill feed would be of sufficient size for testing at one of the mining school or commercial laboratories.

Screen analyses of plant rod mill and laboratory test feeds and products should be made on calibrated Tyler testing sieves.[1]

The surface areas, by Bond,[2] Tyler,[3] or Coghill[4] methods, are computed from the corrected screen analyses.

The calculation of Bond surface, according to the method described in the paper,[2] can be

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[1-5] See references above.

simplified considerably by the construction of a chart on which is plotted intercept K, distribution line slope m, and surface area, each in proper relation to the other. Copies of this chart can be supplied by the writer to interested operators.

With the above information, the power consumption of the large mill (preferably taken with an integrating wattmeter), and its tonnage rate, a calculation of comparative efficiency can be made.

The new surface made in grinding is determined from the computed surface areas, both in the plant mill and in the laboratory test. For example, if we take the case of Tennessee Copper Co., Isabella Mill, Plant Code No. B1, as given in the paper under discussion, we find from Tables 3 and 4:

	Plant Mill	Laboratory Test
Bond surface, mill discharge...	22.00	24.30
Bond surface, mill feed........	5.29	5.50
New Bond surface made....	16.71	18.80

Convert the new Bond surface from square meters per 100 cc solid, as given above, to square meters per kilowatt hour by the following formula:

$$(\text{Bond surface}) \times \frac{9072}{\text{specific gravity}} \times \frac{(\text{tons per hour mill capacity})}{(\text{kw reading for mill})}$$

where Bond surface is given in square meters per 100 cc solid, and 9072 is the number of grams per 2000 lb ton.

For Tennessee Copper B-1, plant mill:

$$\text{Square meters per kw-hr} = 16.71 \times \frac{9072}{4.63} \times \frac{43.8}{128.3} = 11{,}190 \, \frac{\text{sq m}}{\text{kw-hr}}$$

Since there are 3,600,000 joules per kilowatt-hour, the energy in terms of joules per square meter new surface will be:

$$3{,}600{,}000 \div 11{,}190 = 322 \, \frac{\text{joules}}{\text{sq m}} \quad \text{(for plant rod mill)}$$

Now convert the new Bond surface made in the laboratory grindability test from square meters per 100 cc solid to joules per square meter, basing the calculation on an energy input of 83.5 joules per revolution of the laboratory rod mill. The formula is:

$$\frac{83.5 \times 100}{(\text{Bond surface}) \dfrac{(\text{gross grindability})}{(\text{specific gravity})}}$$

where Bond surface is given in square meters per 100 cc solid, and gross grindability is the rate of feeding the laboratory rod mill in terms of grams per revolution of the mill. It is also equal to net grindability divided by the percent /100 of plus 14 mesh material in the laboratory rod-mill feed.

For Tennessee Copper B1, the energy in terms of joules per square meter new surface in the laboratory rod mill is:

$$\frac{8.35 \times 100}{18.80 \times \dfrac{(21.1)}{(4.63)}} = 97.4 \text{ joules per square meter new surface}$$

The ratio of joules per square meter new surface (laboratory rod mill) to joules per square meter new surface (plant rod mill) times 100 equals the relative percentage of power efficiency of the plant rod mill For Tennessee Copper B1, this is:

$$\frac{97.4}{322} \times 100 = 30.2 \text{ pct relative surface area (Bond) efficiency}$$

Similar calculations may be made using Tyler or Coghill surface figures. These figures are based on surface units per unit of weight, and consequently no conversion from a volumetric to a weight basis is required.

To determine relative efficiency of grinding through a definite size, rather than on the surface basis covered by the above calculation, a slightly different procedure must be followed. In this case, determine from the plotted screen analysis the micron size at which 80 pct of the plant rod-mill product passes. Determine the percentage in the plant rod-mill feed passing this micron size, and subtract this percentage of the feed from 80 pct to obtain net production of the rod mill at this micron size, which we choose to designate as "product size," or "P.S." The kilowatt hours per ton utilized by the rod mill divided by the percent net product size made and multiplied by 100 equals kilowatt-hour per net ton of product size.

In the case of Tennessee Copper B1, these figures would be: 80 pct of the plant rod mill product passes 400 microns; 15.8 pct of the plant rod-mill feed is minus 400 microns. The plant rod mill actually produces 64.2 pct minus 400 microns. The plant rod mill uses 2.93 kw-hr per ton and 2.93 kw-hr per ton ÷ 0.642 = 4.57 kw-hr per net ton P.S.

Next determine what the laboratory rod-mill grindability would be at the product size of 400 microns by plotting on the 14 mesh or 1168 micron ordinate (on the same sheet on which the screen analysis of laboratory rod-mill product is plotted) the 14 mesh net grindability. Draw a line through this point parallel to the percentage passing line of the plotted product, and where this new line intersects the 400 micron ordinate read the abscissa value, which is the equivalent grindability at 400 microns or "product size." For Tennessee Copper B1 this is 10.1 g per revolution.

One hundred divided by the product of this product size grindability and the kilowatt hours per net ton at product size equals "Relative P.S. Efficiency," or efficiency of grinding through this mesh size. For Tennessee Copper B1, this would be:

$$\frac{100}{(\text{grindability at P.S.})\ (\text{kw-hr/net ton P.S.})}$$

or

$$\frac{100}{(10.1)(4.57)} = 2.16 \text{ Relative P.S. Efficiency}$$

Generally similar calculations may be developed for determination of relative efficiency at any given mesh.

To compare surface area efficiency with product size efficiency, multiply the latter figure by the ratio of the average of all surface area efficiencies to the average of all product size efficiencies.

The averages of both these quantities will be equal, and in any installation the efficiency of grinding on the basis of surface area produced can be compared directly with the efficiency of grinding to pass product size. In the example B1 cited the mill is much more efficient (50.3 pct) in grinding to pass 400 microns than it is in producing new surface (30.2 pct).

DISCUSSION

(*C. H. Benedict presiding*)

HARLOWE HARDINGE*—Mr. C. H. Benedict, in requesting me to discuss rod mills, has asked for a "large order." The subject cannot be covered in a few words. The compilation of the data given here, and that which is being accumulated in addition, is of real value to the operator and those covering the subject of grinding, in general. I say, "grinding in general," as the rod mill is only one step in the process of comminution of materials. Unless we look at the picture as a whole, we lose sight of the benefits that can be derived from modifications of many of our preconceived notions, prejudices or actual experiences.

Those of us who have been in the grinding game since the advent of the short ball mill as differentiated from open circuit grinding in a relatively long tube mill know that many changes have taken place. We have become influenced by certain results and have gone to extremes at times. When the rod mill was first introduced it was thought by some to be the panacea of all our troubles. Plants were installed and very large ones, too, where the rod mill was even operated in series—primary rod mills, followed by regrind-rod mills. The results compared with the ball mill, particularly, in the regrinding circuit, showed the rod mill was in the wrong field. It got a "black eye" for a time; there were a few staunch advocates who stuck to the mill for use in certain fields and, thanks to them, it did not die. Today, with so much history behind us, the rod mill has been established as a very satisfactory grinding tool in what might be considered the midpoint of the grinding circuit. It is very definitely a competitor of rolls, it is encroaching on the field of the reduction crusher, particularly when handling highly abrasive or sticky ores, and is a direct competitor of the ball mill in the coarse grinding stage, with distinct advantages in its favor where this stage is to be run in open circuit.

The rod mill even has a field in finer grinding in closed circuit with a classifier but the applications are special. The rod mill seems to show up best in larger operations; that is, tonnages in excess of 500 or 700 tons per day where there is opportunity to divide the grind into two or more stages; the rod mill taking the stage from about ¾ in. to 10 mesh, usually in open circuit.

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* Hardinge Co., York, Pa.

B. H. IRWIN*—If large diameter rod mills are more efficient than small diameter mills, there are possible power savings to be had by using a large diameter rod mill instead of several smaller mills.

Large diameter ball mills appear to be more efficient than small in the power required to produce material of a given mesh and much available data bears this out. However the data to substantiate such claims on rod mills has not been available.

Fortunately the data in this paper permits a practical examination of this efficiency problem as concerns the 6 and 9 ft rod mills.

If the kw-hr per net ton is calculated for the meshes 14, 20, 28, 35, 48, 65, 100, 150 and 200, for those mills 6, 6½, 9 and 9½ ft in diameter; and the respective grindabilities are taken from a plot described in the paper in the fifth paragraph of the section on Relative Grinding Efficiencies; the product of the power figures and their respective grindabilities gives the power requirement equivalent to unit grindability in kw-hr per net ton at unit grindability.

The average of the kw-hr per net ton at unit grindability for the 6 and 6½ ft mills and the 9 and 9½ ft mills for each mesh is obtained. The ratio of the difference of the 6 ft average and the 9 ft average to the 9 ft average expressed as a percentage gives the percentage power ratio of the 9 ft mill over the 6 ft mill for each mesh size.

The product of the percentage power ratio and the average net percent produced (percent passing mesh size in product less that in feed) for each mesh gives the relative efficiency of the 9 ft over the 6 ft rod mill for each mesh.

The average relative efficiency of a 9 ft rod mill over a 6 ft mill becomes 11 pct. The resulting data is tabulated below. A plot of the mesh against relative efficiency will show the leveling off of the efficiency at 35 mesh and finer which is well recognized to be nearing the operating extreme of rod mills.

Mesh	Power Ratio, Per Cent	Net Per Cent Produced	Relative Efficiency, Per Cent
14	61	59.5	36
20	39	58.2	23
28	17	54.8	9
35	5	48.4	2
48	4	41.0	2
65	8	33.5	3
100	26	26.2	7
150	36	21.4	8
200	47	16.7	8
Average			11

To completely answer the efficiency problem as concerns a particular installation involves a study as above, other factors to be considered include liner shape and contour, charge segregation, mill shape, electrical and mechanical efficiency. Several of these studies are well established, others are formidable and conclusive evidence will take some time to obtain.

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Standard Grindability Tests Tabulated

By Fred C. Bond,* Member AIME

(New York Meeting, March 1947)

Many grindability tests have been made in the Allis-Chalmers laboratory since the tabulated results were last published.[1] The lists of standard ball-mill and standard rod-mill grindability tests are brought up to date in this paper, and considerable new material is added. This includes a tabulation of open-circuit grinding tests, with impact and compression-crushing tests.

The standard ball-mill grindability tests (Table 1) were made in a standard laboratory ball mill, 12 by 12 in. inside, running at 70 rpm, with a charge of 285 iron balls ranging from 1½ to ¾ in. in diameter and weighing 20,125 grams. Seven hundred cubic centimeters of minus 6-mesh stage-crushed dry feed was used, and the circulating load was maintained constant at 250 pct by adjusting the number of mill revolutions for each grinding period. Tests were made at all mesh sizes from 28 to 200, and the grindability is the number of net grams of screen undersize produced per revolution. The table includes 479 tests. It has been calculated that the ball mill does 52.3 joules of useful work on the ore per revolution[2] with 93 joules input to the mill.

The standard rod-mill grindability tests (Table 2) were made in a laboratory tilting rod mill, 12 in. in diameter by 24 in. long inside, with a wave-type lining, running at 46 rpm. The grinding charge consisted of six 1¼-in. diameter and two 1¾-in. diameter steel rods, all 21 in. long and weighing 33,380 grams. The mill was rotated level for eight revolutions, then tilted up 5° for one revolution, down 5° for one revolution, and returned to level for eight revolutions. It was discharged by tilting downward at 45° for 30 revolutions.

Feed consisted of 1250 cc of minus ½-in. stone and the circulating load was maintained constant at 100 pct by adjusting the number of mill revolutions for each grinding period. Tests were made at all mesh sizes from 3 to 65, and the rod-mill grindability is the number of net grams of screen undersize produced per revolution. The table includes 105 tests.

The impact and compression-crushing tests (Table 3) are given in a table that is an extension of a previous publication[3] in which the method of testing was described in detail. Two equal hammers weighing about 30 lb each were arranged to strike simultaneous blows on opposite sides of the test specimen. Broken pieces of rock passing a 3-in. square and retained on a 2-in. square were set up with the smallest dimension between the hammers, and this distance was measured.

The height of the hammers was increased slightly before each blow until the specimen broke, and the foot-pounds per inch of thickness was calculated. Ten pieces

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[1] References are on next page.

were broken of each ore tested, and both the average and maximum impact crushing strength in foot-pounds per inch are listed.

The compressive strength in pounds per square inch was measured and is listed for many of the ores, determined by breaking approximately four one-inch cubes under slow compression.

The list includes 106 impact tests.

The open-circuit grindability tests (Table 4) have not been published previously. They are concerned with open-circuit wet and dry grinding in two stages. In the first stage 25 lb dry weight of material crushed to minus ½-in. was placed in a 10-in. ball mill of 16½-in. inside diameter, having a smooth lining and running at 44 rpm, and ground with 60 lb of 2-in. iron balls. Water was added if the grinding was to be wet. At intervals of 250 revolutions the mill was stopped, a 250-gram sample was removed. This was screen-analyzed and returned to the mill after the next 250-revolution period, when another sample was removed. This was continued until the charge contained 95 pct or more passing 10-mesh, when the entire charge was removed and dried if necessary.

In the second stage of the test 5 lb of the first mill product was placed in a Abbe Trojan porcelain jar mill with 16 lb of ⅞-in. diameter iron balls, running at 66 rpm. Water was added if the grind was to be wet. Samples weighing 50 grams were removed at intervals of 2500 revolutions, screen-analyzed, and returned to the mill when the next sample was removed. The grind was continued until 92 pct or more passed 200-mesh.

The screen analysis of each stage was plotted separately on linear graph paper. The percentage passing each mesh size was plotted against the number of revolutions. The grinding index for the first stage is the average of the number of revolutions required to grind from 25 to 95 pct passing 10-mesh and from 20 to 85 pct passing 20-mesh.

The grinding index for the second stage is the average of the number of revolutions required to grind from 60 to 99 pct passing 100-mesh and from 50 to 92 pct passing 200-mesh. The table is plotted in the order of increasing values of the second-stage index number, or the order of increasing resistance to open-circuit fine grinding. Any closed-circuit standard ball-mill grindability tests at 200-mesh are included in the tabulation, as are impact and compression-crushing tests.

The values of the second-stage index number vary widely, from the wet grinding of very soft portland cement mixtures and limestones to the dry grinding of hard cement clinkers. The limestones and portland cement mixtures included in the hard-grinding lower half of the table are the mixtures that coat the balls and mill lining, and so are extremely difficult to grind fine.

These open-circuit grinding tests have been made for more than 30 years at the Allis-Chalmers laboratory, but the grinding indices have been calculated only for tests made within the last two or three years. Most of the materials tested in open-circuit are portland cement raw materials, clinker, or limestone. The table lists 49 materials.

REFERENCES

1. F. C. Bond and W. L. Maxson: Standard Grindability Tests and Calculations. *Trans.* A.I.M.E. (1943) **153**, 362.
2. F. C. Bond and W. L. Maxson: Grindability and Grinding Characteristics of Ores. *Trans.* A.I.M.E. (1939) **134**, 296.
3. F. C. Bond: Crushing Tests by Pressure and Impact. *Trans.* (1946) **169**, 58: A.I.M.E. *Mining Tech.* (Jan. 1946—T.P. 1895).

TABLE 1—*Standard Ball-mill Grindability Tests*

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT 28-MESH							
			−28M.				
Gold	Bernheim	S. Rhodesia	1.26*	27.9	1.77		477
Gold	Wright-Hargreaves	Ontario	1.338*	34.2	1.70	2.68	406
Gravel	Pacific Coast Aggregates	California	1.58*	25.2	1.53		324
Copper	Amygdaloid, C. R.	Michigan	1.76*	31.0	1.87	2.93	1000
Gold	Portland	Colorado	1.966*	27.2	1.63	2.69	
Gold	Kerr-Addison	Ontario	2.203*	32.9	1.81		799
Gold	La Luz	Nicaraugua	2.48*	30.0	1.79		932
Copper	New Cornelia	Ajo, Ariz.	2.50*	30.4	1.63	2.68	684
Gold	Little Long Lac	Ontario	2.726*	40.2	1.69	2.64	570
Gold ore	Malartic Goldfields, Ltd.	Halet, Quebec	2.86	33.5	1.74		1592
Gold	San Luis	Mexico	2.905*	20.5	1.67		730
Copper	Shale—White Pine	Michigan	3.08*	33.3	1.79	2.97	1000
Gold	Benquet Consolidated Gold Mines	Baguio, P. I.	3.12	31.4	1.69	2.66	550-B
Copper	Quincy	Michigan	3.12*	21.0	1.98		1036
Gold	Rand-Springs Mine	S. Africa	3.22*	23.0	1.78	2.71	504
Gold	Kerr-Addison	Ontario	3.365*	32.7	1.83		799
Copper	Miami Ore	Miami, Ariz.	3.37*	26.9	1.64	2.69	
Quartz	Can. Silica Products	Quebec	3.58	11.7	1.56	2.65	398
Copper	Castle Dome Ore	Miami, Ariz.	3.655*	24.8	1.685		1042
Copper	Utah Copper—Arthur	Utah	3.91*†	26.5	1.715		938
Gold	Homestake	S. Dakota	3.95*	35.5	1.91	3.12	
D.-L.[a] sinter	Eagle Picher	Illinois	4.01	21.1	2.25		341
Copper	Anaconda Copper Co.	Anaconda, Mont.	4.14	20.4	1.93		1477-A
Silver	Hualgayoc	Peru	4.27	25.2	2.45		346
Copper	Sandstone—Vein Rock	Michigan	4.403*†	20.7	1.83	2.68	1000
Copper	Anaconda	Montana	4.44*	34.3	2.18	3.23	910
Quartz	Pure Crystallized	California	4.55*	17.0	1.68	2.65	
Magnetite	M. A. Hanna, Clifton	New York	4.57†	14.8	2.76		1022
Iron	U. of Minn. Oglebay, Norton & Co.	Babbitt, Minn.	5.05	24.5	2.40		1167-A
Cassiterite	Vulcan Detinning	New Jersey	5.09	23.1	3.60		558
Gold	East Malartic	Quebec	5.42*	37.8	1.87	2.79	779
Syenite	Can. Nepheline Syenite	Ontario	5.51†	16.7	1.63	2.73	663
Arsenic-gold	Getchell Mines, Inc.	Red House, Nev.	5.58	20.4	1.70		1241-A
Magnetite	Republic Steel, Chateaugay	New York	6.265*	8.8	2.98		868
Copper	Morenci	Arizona	6.34*†	24.8	1.57	2.63	913
Copper	Nevada Consolidated Copper Co.	Chino Mines, Hurley, N. M.	6.44	32.6	1.83		1060-A
Copper	Sandstone—White Pine	Michigan	8.63*	13.8	1.70	2.635	1000
Phosphate	Charleston Min. Co.	Tennessee	8.87†	14.3	1.86		333
Gar sand	Garco Products Co.	Ford City, Pa.	10.33	43.4	1.25		1181-A
Roasted	Sullivan Min. Co.	Idaho	15.33*	38.9	1.93		740
Gold	Red Cross Min. Co.	California	20.9	26.7	1.56		331
Avg.			4.83	26.32			
TESTS AT 35-MESH							
			−35M.				
Cast iron	Du Pont	Delaware	0.062	8.6	3.12		325
Graphite	Alcoa	Tennessee	0.24	25.4	1.01		937
Cast iron	Crucible Steel	Pittsburgh	0.611	2.6	0.94		436
Copper	Conglomerate, C. & H.	Michigan	1.231*	33.5	1.69		771
Gravel	Pac. Coast Aggregates	California	1.28*	28.2	1.53		324
Copper	Amygdaloid, C. R.	Michigan	1.53*†	36.5	1.87	2.93	1000
Copper	Amygdaloid, C. & H.	Michigan	1.82*	30.3	1.75		771
Silica	Ottawa Silica Co.	Ottawa, Illinois	1.88	7.8		2.65	1727
Gold	La Luz	Nicaragua	2.08*	36.0	1.79		932
Gold	Kerr-Addison	Ontario	2.17*	37.3	1.83		799
Copper	Tailings—Quincy	Michigan	2.18*	28.9	1.98		1036
Copper	New Cornelia	Ajo, Ariz.	2.31*	34.7	1.63	2.68	684
Gold	Matartic Goldfields Co.	Halet, Quebec	2.31	38.8	1.74		1592
Gold	Little Long Lac	Ontario	2.35*	32.6	1.69	2.64	570

[a] Dwight-Lloyd.

* Standard ball-mill tests made at other mesh sizes.

† Standard rod-mill tests made on this ore.

TABLE I—(*Continued*)

TESTS AT 35-MESH—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
			−35M.				
Gold	Benguet	P.I.	2.38*	35.6	1.69	2.66	550
Gold	Rand-Springs Mine	So. Africa	2.404*†	27.7	1.78	2.71	504
Gold	San Luis	Mexico	2.495*	25.7	1.67		730
Copper	Shale—White Pine	Michigan	2.66*	38.2	1.79	2.97	1000
Gold	Kerr-Addison	Ontario	2.95*	37.6	1.81		799
Iron	Bowring	Morocco	2.97*	25.0	1.86		288
Copper	Castle Dome Ore	Miami, Ariz.	2.99*	29.5	1.68		1042
Iron	Bowring	Spain	3.03*	20.9	2.12		291
Copper	Anaconda	Montana	3.31*	29.3	2.18	3.23	910
Copper	Anaconda Copper Mining Co.	Anaconda, Mont.	3.42	24.4	1.93		1477-A
Copper	Utah Copper—Arthur	Utah	3.91*†	30.0	1.71		938
Arsenic-gold	Getchell Mines, Inc.	Red House, Nev.	3.50	38.2	1.70		1241
Iron	U. of Minn. Oglebay Norton & Co.	Babbitt, Minn.	3.96	30.0	2.40		1167-A
Copper	Sandstone—Vein Rock	Michigan	4.05*†	28.8	1.83	2.68	1000
Copper	Nevada Consolidated Copper Co.	Chino Mines, Hurley, N. M.	4.11	30.7	1.83		1060
Copper	Morenci	Arizona	4.50*†	27.5	1.57	2.63	913
Copper	Sandstone—White Pine	Michigan	4.51*	20.4	1.70	2.64	1000
Iron	Republic Steel, Chateaugay	New York	4.70*	12.3	2.98		868
Feldspar	Consolidated Feldspar Corp.	Erwin, Tenn.	4.90	12.3	1.68		1103-B
Iron	Alan Wood Steel, Croton	New York	5.35*	16.2	2.18		931
Clinker	Cobrecite	Michigan	7.10*	12.6	0.76		830
Gold	Mineral Mining Corp.	Kershaw, S. Carolina	7.94	54.6	1.436		782-A
Roasted	Sullivan Mines	Idaho	8.04*	40.8	1.93		740
Avg.			3.16	27.83			

TESTS AT 48-MESH

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity Apparent	Specific Gravity True	No.
			−48M.				
Grinding-wheel chips	Allegheny-Ludlum Corp.	Waterville, N. Y.	.173	17.8	1.75		1089-A
Petr. coke	Alcoa	Tennessee	0.304	26.9	0.97	1.78	700
Scrap emery	Milwaukee Steel	Wisconsin	0.408*†	40.3	1.93		989
Zinc	National Zinc	Oklahoma	0.572	23.7	0.97		161
Emery-sandstone	Milwaukee Steel Co. & Copper Range Co.	Milwaukee, Wis. & Painesdale	.779	39.8	1.88		1015-B
Emery-copper	Milwaukee Steel Co. & Phelps Dodge Co.	Milwaukee & Morenci, Ariz.	.780	40.2	1.75		1015-C
Refractory	Laclede Christy	Illinois	0.86*	24.4	1.65		941
Gold	Bernheim	S. Rhodesia	0.87*	42.7	1.77		477
Flint	Tri-State Flint	Missouri	0.90*	29.4	1.54		440
Gold	Wright-Hargreaves	Ontario	1.24*	44.0	1.70	2.68	406
Copper	Amygdaloid, C. R.	Michigan	1.37*†	42.3	1.87	2.93	1000
Gold	Kelowna Expl.	B. Columbia	1.37*	47.5	2.09	3.22	676
Gold	Noranda	Quebec	1.47*	44.9	1.85	2.86	
Gold	Portland	Colorado	1.64*	36.7	1.63	2.69	
Brick mix	Carnegie Steel	Illinois	1.82	24.5	1.54		907
Gold	Little Long Lac	Ontario	1.826*	41.4	1.693	2.64	570
Gold	La Luz	Nicaragua	1.855*	41.0	1.79		932
Copper	Tailings—Quincy	Michigan	1.950*	35.7	1.98		1036
Gold	A. O. Smith Corp., R. M.	Nevada	1.96	34.9	1.61		646
Firebrick	Carnegie Steel	Illinois	1.98	24.5	1.49		907
Gold	Rand-Springs Mine	S. Africa	1.985*	35.9	1.78	2.71	504
Silver	Real Del Monte	Pachuca	2.00*	46.1	1.63		881
Copper	New Cornelia	Ajo, Ariz.	2.10*	40.0	1.63	2.68	684
Gold	San Luis	Mexico	2.123*	27.7	1.67		730
Gold	Benquet Consolidated Gold Mines	Baguio, P.I.	2.19	38.0	1.69	2.66	550
Gold	Malartic Goldfields Co.	Halet, Quebec	2.25	42.0	1.74		1592
Gold	San Fernando	Mexico	2.28	34.6	1.64		255
Copper	Shale-White Pine	Michigan	2.296*	45.4	1.79	2.97	1000
Gold	San Fernando	Mexico	2.31	35.8	1.56		255

* Standard ball-mill tests made at other mesh sizes.
† Standard rod-mill tests made on this ore.

TABLE I—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
		TESTS AT 48-MESH—(*Continued*)					
			−48M				
Gold	Saramarca	Peru	2.39	32.4	1.82		547
Garnet	K. T. Felder	Georgia	2.40	36.5	1.78		269
Phosphate	Al-Ke-Me Fertilizer	Brazil	2.41*	32.9	1.39		252
Copper	Castle Dome Ore	Miami, Ariz.	2.52*	36.1	1.68		1042
Copper	Miami Ore	Miami, Ariz.	2.61	38.2	1.64	2.69	
Lead-zinc	Montecatini	Italy	2.69*	31.4	2.32		469
Quartz	Pure Crystallized	California	2.72*	27.1	1.68	2.65	
Copper	Anaconda	Montana	2.75*	34.4	2.18	3.23	910
Gold	Homestake	S. Dakota	2.95*	45.1	1.91	3.12	
Gold	Picacho Min. Co.	Arizona	2.956*	31.4	1.52		880
Copper	Sherritt Gordon	Manitoba	2.97	31.2	1.86		
Copper	Cons. Copper Co.	Nevada	3.00	35.0	1.54		188
Copper ore	Anaconda Copper Mining Co.	Anaconda, Mont.	3.13	27.7	1.93		1477-A
Copper	Sandstone—White Pine	Michigan	3.05*	28.5	1.70	2.64	1000
Spodumene	Solvay Process	New York	3.15	23.1	1.73		1073
Copper	Sandstone—Vein Rock	Michigan	3.15*†	28.7	1.83	2.68	1000
Rock	Emsco Refractories Co.	Lehi, Utah	3.20	22.0	1.67		1255-A
Copper	Utah Copper—Arthur	Utah	2.23*†	36.9	1.72		938
Copper	Morenci	Arizona	3.33*†	33.8	1.57	2.63	913
Iron	U. of Minn. Oglebay, Norton & Co.	Babbitt, Minn.	3.42	38.2	2.40		1167
Shale	St. Lawrence Brick	Quebec	3.50*	40.0	1.59		279
Copper	Nevada Consolidated Copper Co.	Chino Mines, Hurley, N. M.	3.66	39.6	1.83		1060-A
Fontana rock	T. V. A. Fontana Dam	Knoxville, Tenn.	3.69	32.1		2.65	1750
Gold	East Malartic	Quebec	3.698*	37.8	1.87	2.80	779
Copper	Silver Bell	Arizona	3.74*	37.1	1.69		1050
Gold	C. M. & R.—Golden Rose	B. Columbia	3.77*	40.5	2.10		864
Gold	S. A. Devel. Co.	Ecuador	4.08*	38.7	2.67	3.81	752
Iron	Alan Wood Steel, Croton	New Jersey	4.34*	21.6	2.18		931
Isabella mill	Tennessee Copper Co.	Copperhill, Tenn.	4.72	34.6		4.63	1750
Copper	Tennessee Copper	Tennessee	4.80*	30.2	2.08		263
Quartzite	Emsco Refractories Co.	Lehi, Utah	4.82		1.54		1178-A
Copper	Cons. Copper Co.	Nevada	5.32	38.2	1.64		189
Phosphate	Moinho Santista-Brazil	Brazil	5.42	16.7		3.38	1732-B
Magnesium	Basic Refr. Inc.	Ohio	5.465*	48.4	2.134	2.927	1033
Lead-zinc	St. Joseph Lead Co.	Emeryville, N. Y.	5.58	21.5	2.10	3.46	1657
Al sinter	Aluminum Co. of Canada	Quebec	5.70*†	37.4	0.83		1080
Copper	Bingham Canyon, U. S. M.	Utah	5.90*	30.7	2.68	4.65	
Iron	Inland Steel, B. R. F.	Wisconsin	6.04*	30.2	2.11		837
Zinc	Farrey Min. & Mill. Co.	Illinois	6.74		1.74		184
London mill	Tennessee Copper Co.	Copperhill, Tenn.	8.33	28.6		4.52	1750
Graphite	C. A. Condon	Alabama	9.26	24.5	1.35		287
Iron sinter	Sloss-Sheffield Steel & Iron Co.	Birmingham, Ala.	10.46	12.1		1.86	1158-A
Tripoli	Western Minerals	Kansas	12.33†	75.5	1.14		883
Avg.			3.23	34.50			
		TESTS AT 65-MESH					
			−65M.				
Cast iron	Miami Copper	Arizona	0.044	29.3	3.51	7.07	372
Chromium metal	Electro Met. Co.	New York	0.313*†	10.9	3.91		1008
Emery	Am. Emery Wheel Wrks.	Rhode Island	0.459†	51.2	2.47	3.89	410
Copper	Amygdaloid, C. R.	Michigan	1.180*†	50.5	1.87	2.93	1000
Gold	Kelowna Expl.	B. Columbia	1.24*	54.4	2.09	3.22	676
Gold	Little Long Lac	Ontario	1.49*	50.8	1.69	2.64	570
Silver	Real Del Monte	Pachuca	1.53*	45.6	1.63		881
Gold	La Luz	Nicaragua	1.55*	50.6	1.79		932
Copper	Tailings—Quincy	Michigan	1.575*	47.0	1.98		1036
Copper	Britannia Min. & Sm.	B. Columbia	1.577	39.1	1.71		326
Gold	A. O. Smith	Wisconsin	1.583	54.2	1.60		587
Coal	Illinois Zinc	Illinois	1.61	47.8	1.69	2.80	376

* Standard ball-mill tests made at other mesh sizes.
† Standard rod-mill tests made on this ore.

TABLE I—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT 65-MESH—(*Continued*)							
			−65M.				
Copper.........	Sandstone—White Pine	Michigan	1.623*	42.2	1.70	2.64	1000
Sea shells.......	Calizos	Chile	1.637	43.8	1.52		967
Gold...........	Rand-Springs Mine	S. Africa	1.717*	44.7	1.785	2.71	504
Gold...........	Madsen Red Lake	Ontario	1.730*	40.5	1.70	2.65	626
Gold...........	G. E. Smith	Oregon	1.85	37.9	1.60		321
Granite.........	Picacho	Arizona	1.877	39.3	1.53	2.64	517
Gold...........	Spring Hill	Calif.	1.903	45.9	1.70		538
Copper.........	Shale—White Pine	Michigan	1.930*	54.2	1.79	2.97	1000
Iron............	Bowring	Morocco	1.94*	36.8	1.86		288
Copper.........	New Cornelia	Ajo, Ariz.	2.022*	45.0	1.63	2.68	684
Copper.........	Castle Dome	Miami, Ariz.	2.023*	44.9	1.685		1042
Gold...........	Malartic Goldfields, Ltd.	Halet, Quebec	2.04	51.6	1.74		1592
Copper.........	Chelan Copper	Washington	2.07	44.5	1.64		328
Copper.........	Cons. Copper Co.	Nevada	2.10	43.7			278
Copper.........	Cyprus Mines	Cyprus	2.110*	43.4	2.69		432
Copper.........	Stadacona Rouyn	Quebec	2.157	53.2	1.89		561
Serpentine......	G. E. Baker Co.	Pennsylvania	2.246	64.2	1.70		1051
Phosphate.......	Ipanema Plant	Brazil	2.255*	42.8	2.36		352
Gold...........	Picacho Min. Co.	Arizona	2.26*	46.4			880
Gold...........	Cline Lake	Ontario	2.29*	54.7	1.78		745
Manganese......	C. L. Walfred	Minnesota	2.30*	50.7	1.64		867
Gold...........	Santa Maria del Oro	Mexico	2.31*	33.3	2.08	3.10	574
Gold...........	LaLuz	Nicaragua	2.32*	44.7	1.79		932
Iron............	Moose Mountain	Ontario	2.33*†	53.9	2.44	3.42	756
Arsenic-gold.....	Getchell Mines, Inc.	Red House, Nev.	2.33	51.4	1.70		1241-A
Gold...........	The Merrill Company	San Francisco, Calif.	2.34	36.9			1722
Copper.........	Sandstone, C. R.	Michigan	2.34*†	37.4	1.83	2.68	1000
Lead-zinc.......	Montecatini	Italy	2.39*	42.6	2.32		469
Gold...........	Parcoy	Peru	2.393*	31.3	2.26		567
Gold...........	Minnesota Mines	Colorado	2.420*	43.0	1.85	2.81	637
Copper.........	Anaconda	Montana	2.435*	41.2	2.18	3.23	910
Copper.........	Morenci	Arizona	2.52*†	44.9	1.57	2.63	913
Copper.........	Anaconda Copper Co.	Anaconda, Mont.	2.53	34.0	1.93		1477
Gold...........	Homestake	S. Dakota	2.55*	53.7	1.91	3.12	
Iron............	U. of Minn., Oglebay, Norton & Co.	Babbitt, Minn.	2.59	46.9	2.40		1167-A
Gold...........	Ipo Mine	P.I.	2.61*	37.2	1.68		329
Copper.........	Utah Copper—Arthur	Utah	2.65*†	46.6	1.715		938
Gold...........	A. O. Smith	Wisconsin	2.675	55.0	1.60		587
Copper.........	Nevada Consolidated Copper Co.	Chino Mines, Hurley, N. M.	2.68	41.2	1.83		1060-A
Salt............	M. D. P. d'Alsace	France	2.685*†	22.9	1.32		726
Tungsten.......	Nevada-Mass. Co.	Nevada	2.69†	37.0	1.71		942
Iron oxide sinter	Combustion Engineering Co.	Chattanooga, Tenn.	2.79	30.5	2.15		1201-A
Copper.........	Silver Bell	Arizona	2.80*	46.0	1.68		1050
Iron............	Alan Wood Steel, Croton	New Jersey	2.86*	31.2	2.18		931
Quartzite.......	Ensco Refractories Co.	Lehi, Utah	2.88	21.6	1.45		1178-A
Gold...........	S. A. Devel. Co.	Ecuador	3.02*	50.8	2.67	3.81	752
Limestone.......	H. J. Kaiser Co.	California	3.123	45.5	1.66		877
Gold...........	East Malartic	Quebec	3.14*	58.1	1.87	2.79	779
Copper.........	Tennessee Copper	Tennessee	3.29*	40.9	2.08		263
Magnesite.......	N. W. Magnesite Co.	Washington	3.370*	38.6	3.09		1031
Copper.........	Cons. Copper	Nevada	3.51	46.0	1.64		276
Slag............	Monsanto Chem.	Alabama	3.81*	17.3	1.90		1045
Iron............	Alan Wood Steel	New Jersey	4.15†	36.6	2.41	3.40	914
Fe-Si, 25-75.....	Electro-Met. Co.	New York	4.17*	33.9	1.94		1067
Al sinter........	Aluminum Co. of Canada	Quebec	4.23*†	36.0	0.83		1080
Clay............	H. J. Kaiser Co.	California	4.235	32.7	1.36		877
Magnesium.....	Basic Magnesium	Nevada	4.270*	54.0	2.13	2.93	1033
Fe-Mn alloy.....	Champion Rivet Co.	Ohio	4.523	35.0	4.63		442
Iron ore sinter...	Sloss-Sheffield Steel & Iron Co. National Lead Co.	Birmingham, Ala.	4.92	22.6	1.86		1158-A
Fluorspar.......	Aluminum Ore Co.	Illinois	5.700	35.0	1.88	2.98	619
Clinker.........	Cobrecite	Michigan	6.82	17.5	0.76		830
Pyrite..........	St. Joseph—Conc.	New York	21.7	30.7	2.99		847
Bauxite.........	Swann and Co.	Alabama	51.2	23.3	1.08		738
Avg............			3.42	41.6			

* Standard ball-mill tests made at other mesh sizes.
† Standard rod-mill tests made on this ore.

TABLE 1—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT 100-MESH							
			−100M.				
Petr. coke	Alcoa	Tennessee	0.19	56.0	1.01		700
Cr Metal	Electro-Met. Co.	New York	0.2115*†	21.2	3.91		1008
Gold	Chas. Butters	Nicaragua	0.393*	56.7	1.53		338
Flint	Tri-State Flint	Missouri	0.618*	52.6	1.54		440
Gold	Bernheim	S. Rhodesia	0.79*	60.0	1.77		477
Gold	Veraguas Mines	Panama	0.832	71.8	1.75	2.78	498
Copper	Conglom.—C. & H.	Michigan	0.833	63.4	1.69		771
Abrasive	Monsanto Chem.	Alabama	0.858	30.2	2.11	3.91	656
Fire clay	Standard Fuel	Michigan	0.964*	46.8	1.25		835
Gold	Wright-Hargreaves	Ontario	0.98*	61.0	1.70	2.68	406
Copper	Copper Range Co.	Freda, Mich.	.99	61.0	1.91		1599
Gold	Kelowna Expl.	B. Columbia	1.01*	67.5	2.09	3.22	676
Gold	Western Mach. Co.	Calif.	1.040	59.4	1.62		903
Gold	Noranda	Quebec	1.079	69.9	1.85	2.86	
Gold	Little Long Lac	Ontario	1.150*	67.8	1.69	2.64	570
Gold	Wright-Hargreaves	Ontario	1.15*	67.7	1.70	2.68	406
Copper	Amygdaloid—C. & H. Min.	Michigan	1.185*	66.1	1.745		771
Silver	Real Del Monte	Pachuca	1.230*	56.5	1.63		881
Copper	Sandstone—White Pine	Michigan	1.250*	55.6	1.70	2.64	1000
Gold	Portland	Colorado	1.256	56.6	1.63	2.69	
Gold	Amer. Cyanamid	New Jersey	1.288	62.7	1.68	2.61	524
Sp. iron	Ford Motor	Michigan	1.295*	46.8	3.48		1035
Gold	LaLuz	Nicaragua	1.30*	63.4	1.79		932
Gold	Rand-Springs Mine	S. Africa	1.323	62.3	1.78	2.71	504
Gold	W. A. Liddell	Texas	1.355	46.3	1.75	2.59	632
Shale	Minn. Mining & Mg. Co.	Wausau, Wis.	1.34	55.0	1.53		1217
Fire clay	Standard Fuel	Michigan	1.355*	52.4	1.23		835
Phosphate	Al-Ke-Me	Brazil	1.36	61.8	1.39		252
Gold	San Luis	Mexico	1.41*	49.7	1.67		730
Copper	Tailings—Quincy	Michigan	1.413*	59.3	1.98		1036
Gold	E. T. Merritt	Ontario	1.46	55.8	1.64		301
Phosphate	Ipanema Plant	Brazil	1.47*	59.6	2.36		352
Copper	Kanshanshi	Congo	1.49	62.9	1.32		316
Silver	Cia. Min. Carlota	Chile	1.51*	72.0	1.88	3.00	660
Gold	Madsen Red Lake	Ontario	1.515*	53.4	1.70	2.65	626
Quartz	Pure Crystallized	Calif.	1.523*	53.0	1.68	2.65	
Gold	Getchell Mine	Nevada	1.535	59.3	1.45		714
Silver	Tonopah Min.	Nevada	1.54	62.35	1.77		802
Copper	Shale—White Pine	Michigan	1.56*	70.6	1.79	2.97	1000
Copper	New Cornelia	Ajo, Ariz.	1.57*	73.1	1.63	2.68	684
Gold	Santa Maria Del Oro	Mexico	1.575*	45.7	2.08	3.10	574
Copper	Sandstone—Vein Rock	Michigan	1.577*†	56.7	1.83	2.68	1000
Gold	Dahlonega Gold	Georgia	1.58	51.9	1.67		250
Slate	Maine Development Comm.	Augusta, Me.	1.60		1.71		1679
Gold	Sherritt Gordon	Manitoba	1.64	57.0	1.85		206
Gold	Malartic Goldfields Ltd.	Halet, Quebec	1.67	63.3	1.74		1592
Gold	Am. Metals	New Jersey	1.676*	50.0	1.74		808
Manganese	Gen. Manganese Co.	S. Dakota	1.71		1.62		734
Gold	Ziebright Mine	Calif.	1.752	56.4	1.79		505
Copper	Castle Dome	Miami, Ariz.	1.783*	56.1	1.69		1042
Copper	Cyprus Mines	Cyprus	1.786*	60.3	2.69		432
Tin	Tainton Products	Bolivia	1.80	52.1	2.42		905
Gold	F. Viles	Montana	1.81	63.9	1.72		564
Copper	Miami Ore	Miami, Ariz.	1.816*	52.2	1.64	2.69	
Manganese	C. C. Walfred	Minn.	1.83	60.0	1.69		867
Calcium phosphate	Anaconda Copper Mining Co.	Soda Springs, Idaho	1.84	33.5	1.82		1226-A
Copper	Anaconda Copper	Montana	1.85*	54.7	2.18	3.23	910
Gold	Rochester-Plymouth	Nevada	1.860*	63.1	1.83	2.75	518
Arsenic-gold	Getchell Mines, Inc.	Red House, Nevada	1.87	62.4	1.70		1241
Copper	Morenci	Arizona	1.88*†	54.5	1.57	2.63	913
Gold	Cline Lake	Ontario	1.92*	67.3	1.78		745
Gold	Homestake	S. Dakota	1.964*	66.5	1.91	3.12	
Copper	Silver Bell	Arizona	1.98*	59.6	1.68		1050
Gold	Atlantic Gulf & Pacific	P.I.	2.01*	52.9	1.68		329
Copper	Anaconda Copper Co.	Anaconda, Mont.	2.05	46.8	1.93		1477-A

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* Standard ball-mill tests made at other mesh sizes.

† Standard rod-mill tests made on this ore.

TABLE 1—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT 100-MESH—(*Continued*)							
			−100M.				
Iron oxide.......	U. of Minn., Oglebay, Norton & Co.	Babbitt, Minn.	2.10	58.3	2.40		1167-A
Copper.........	Utah Copper—Arthur	Utah	2.15*†	57.8	1.72		938
Slag............	Monsanto Chem.	Alabama	2.16*	31.6	1.90		1045
Tin.............	Pitts. Plate Glass	Mexico	2.226*	55.5	2.92		1018
Phosphate......	Int. Agric. Corp.	Florida	2.260	47.3	1.50		394
Fe-Cr alloy.....	Chromium Min. & Smelt.	Ontario	2.27	47.8	3.97		743
Tin.............	Pitts. Plate Glass	Mexico	2.335*	56.7	3.04		1018
Nickel..........	Falconbridge	Ontario	2.405	59.3	2.29	3.65	371
Magnesite.......	N.W. Magnesite Co.	Washington	2.410*	50.3	3.09		1031
Gold...........	S. Amer. Dev. Co.	Ecuador	2.44*	63.1	2.67	3.81	752
Fontana quartzite	T.V.A. Fontana Dam	Knoxville, Tenn.	2.47	58.5	1.83		1055-D
Gold...........	E. Malartic	Quebec	2.49*	68.0	1.87	2.80	779
Iron............	Marquette Carbonate	Wisconsin	2.50	79.2	2.17		342
Gold...........	C. M. & S.—Golden Rose	B. Columbia	2.52*	58.2	2.10		864
Copper.........	Cons. Copper	Nevada	2.59*	60.2	1.64		276
Iron............	Bowring, N. Y.	Morocco	2.66*	49.4	1.80		315
Iron............	Iron River Falls	Wisconsin	2.72*	58.7	2.39		1023
Iron ore sinter...	Sloss-Sheffield, Steel & Iron Co.	Birmingham, Ala.	2.75	39.9	1.86		1158-A
Fe-Si, 25–75.....	Electro-Met. Co.	New York	2.950*	47.2	1.94		1067
Copper.........	Bingham Canyon	Utah	2.99	42.6	2.68	4.65	
Gold...........	Butte-Highlands	Montana	3.13*	58.3	1.67		861
Iron............	U. of Minn. Oglebay, Norton & Co.	Babbitt, Minn.	3.14	41.2	2.32		1167-B
Salt............	Du Pont	New York	3.48	44.8	1.39		939
Magnesium.....	Basic Refr. Inc.	Ohio	4.125*	61.9	2.13	2.93	1033
Slag............	Victor Chem. Wks.	Tenn.	4.15	48.0	4.05		284
Ferro-phosphorus	Monsanto Chemical Co.	Columbia, Tenn.	29.1	54.6	4.25		1262-A
Avg............			2.06	56.4			
TESTS AT 150-MESH							
			−150M.				
Graphite........	Long Valley—Cons.	New York	0.31*	49.8	0.96		860
Coke...........	Bertelsen Co.	Bridgeville, Pa.	.40	67.7	.802		1595
Copper.........	Sandstone—White Pine	Michigan	0.85*	84.1	1.70	2.64	1000
Gold...........	Kelowna Expl.	B. Columbia	0.86*	77.9	2.09	3.22	676
Quartz..........	Fused Quartz sand	Illinois	1.02	63.5	1.48		854
Flint granules...	Thetford Mines	Quebec	1.04	63.5	1.64		1604
Gold...........	Little Long Lac	Ontario	1.080*	82.6			570
Gold...........	Rand Springs Mine	S. Africa	1.117*	76.7	1.78		504
Gold...........	H. C. Winans	Brazil	1.155	76.8	1.63		474
Pottery.........	John Douglas Co.	Cincinnati, Ohio	1.20	24.5	1.51		1472
Gold...........	Powell-Rouyn	Quebec	1.233*	79.5	1.78		949
Copper.........	Sandstone—White Pine	Michigan	1.249*	73.1	1.78		1000
Gold...........	Bong Mieu	Indo-China	1.31	75.7	2.00	3.07	339
Gold...........	Kerr-Addison	Ontario	1.32*	81.6	1.81		799
Copper.........	Anaconda	Montana	1.36*	75.2	2.18	3.23	910
Gold...........	Malartic Goldfields Ltd.	Halet, Quebec	1.36	78.2	1.74		1592
Gold...........	Minnesota Mines	Colorado	1.368*	77.6	1.85	2.81	637
Copper.........	Morenci	Arizona	1.395*†	77.2	1.57	2.63	913
Copper.........	Anaconda Copper Co.	Anaconda, Mont.	1.40	75.0	1.93		1477
Copper.........	Castle Dome	Miami, Ariz.	1.396*	78.7	1.69		1042
Iron............	Minerals Separation of N. A.		1.41	75.7	1.5		1193-A
Copper.........	Mines De Bor	Yugoslavia	1.41*	84.5	1.91		249
Copper.........	Shale—White Pine	Michigan	1.430*	83.7	1.79	2.97	1000
Gold...........	Can.-Malartic Min. Co.	Quebec	1.445	78.5	1.84		586
Gold...........	Parcoy	Peru	1.495*	72.8	2.26		567
Iron............	U. of Minn. Oglebay, Norton & Co.	Babbitt, Minn.	1.52	79.2	2.40		1167-A
Gold...........	Kerr-Addison	Ontario	1.567*	79.3	1.83		799
Nickel..........	B. C. Nickel Mines	B. Columbia	1.607	71.4	2.47		716

* Standard ball-mill tests made at other mesh sizes.
† Standard rod-mill tests made on this ore.

TABLE 1—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT 150-MESH (*Continued*)							
			−150M.				
Langbeinite.....	Union Potash & Chem.	New Mexico	1.609	77.0	1.75		1006
Copper.........	Utah Copper—Arthur	Utah	1.62*†	77.1	1.72		938
Copper.........	Nevada Consolidated Copper Co.	Chino Mines, Hurley, N. M.	1.65	76.2	1.83		1060-A
Tin.............	Pitts. Plate Glass	Mexico	1.675*	74.6	2.92		1018
Iron............	Moose Mountain	Ontario	1.689*†	81.5	2.44	3.42	756
Gold...........	Buffalo-Ankerite	Ontario	1.705	82.9	1.94	3.22	614
Gold...........	M. A. Smith	Cuba	1.725	78.2	1.86		710
Tin.............	Pitts. Plate Glass	Mexico	1.783*	76.2	3.03		1018
Zinc-lead........	Callahan Zinc-Lead	Idaho	1.81	60.4	2.23		691
Limestone.......	Lawrence Cement Co.	Pennsylvania	1.870*	76.2	1.90		972
Iron............	Iron River Falls	Wisconsin	1.916*	77.3	2.39		1023
Gold...........	Preston East Dome	Ontario	2.01*	86.8	1.78		694
Gold...........	C. M. & S.—Golden Rose	B. Columbia	2.03*	74.1	2.10		864
Iron............	Inland Steel Co.	Wisconsin	2.09*	62.8	2.10		837
Gold...........	E. Malartic	Quebec	2.134*	83.5	1.87	2.80	779
Gold...........	Atienda	Italy	2.84	84.4	1.42	2.56	629
Magnesium.....	Basic Magnesium	Nevada	2.855*	79.5	2.13	2.93	1033
Nickel..........	Nicaro Nickel Co.	Cuba	3.48	73.8	1.13		869
Nickel..........	Nicaro Nickel Co.	Cuba	5.55	86.0	1.29		1075
Avg............			1.62	75.3			
TESTS AT 200-MESH							
			−200M.				
Graphite........	Long Valley—Conc.	New York	0.23*		0.96		860
Silicon carbide...	Electro-Ref.	New York	0.259		1.89		1047
Ferro-silicon.....	Basic Ind. Lab. (Allis-Chalmers)	Milwaukee, Wis.	.263		3.98		1612-A
Flint...........	Tri-State Flint	Missouri	0.491*				440
Gold...........	Bernheim	S. Rhodesia	0.56*		1.77		477
Shale...........	Korite Corp.	Wisconsin	0.59		1.59		677
Copper.........	Granby Consolidated		.59		1.70		1566
Silicon carbide...	Electro-Ref.	New York	0.614		1.25		1047
Titanium.......	Titanium Corp.	Arkansas	0.620		2.36		559
Copper.........	Amygdaloid C. & H. Min.	Michigan	0.627*		1.69		771
Magnesite.......	Harbison-Walker Refractories	Indiana	.65			3.36	1768
Sp. Iron........	Ford Motor Co.	Michigan	0.664*		3.48		1035
Min. Wool......	Mineralite Corp.	Penna.	0.678		0.19		399
Gold...........	San Luis	Mexico	0.688*		1.67		730
Iron............	Mabellite Corp.	New York	0.719		1.74		419
Silver..........	Real Del Monte	Pachuca	0.720*		1.63		881
Copper.........	Sandstone—White Pine	Michigan	0.733*		1.20	2.64	1000
Gold...........	Kelowna Expl.	B. Columbia	0.758*		2.09	3.27	676
Gold...........	Lake Shore Gold Mines Ltd.	Kirkland Lake, Ont.	.79		1.51		1644
Copper.........	Tailings—C&H Min.	Michigan	0.759*		1.75		771
Gold...........	Buffalo Ankerite Gold Mines Ltd.	S. Porcupine, Ont.	.77		1.83		1639
Gold...........	Wright-Hargreaves	Ontario	0.771*		1.70	2.68	406
Gold...........	Cia. Minera Ciclon	Chile	0.788		1.84		719
Clay...........	Sun Oil Co.	Pennsylvania	0.790		0.66	1.09	641
Cement clinker..	Consolidated Cement Corp.	Cement City, Mich.	.81		1.67		1617-B
Gold...........	Carrizalillo	Chile	0.816*		1.74		808
Apatite-phosphate	Bunge Corp.	Moinho Santista, Brazil	.81		1.67		1555-B
Magnetite......	American Cyanamid Corp.	Stanford, Conn.	.82		3.06		1511
Copper.........	Copper Range Co.	Painesdale, Mich.	.842		1.777		1000-H
Shale...........	Minn. Mining & Mfg.	Wausau, Wis.	.858		1.53		1217-A
Copper.........	Noranda	Quebec	0.83*		1.85	2.86	
Gold...........	Rand-Springs Mine	S. Africa	0.859*		1.78	2.71	504
Shale...........	Korite Corp.	Wisconsin	0.86		1.36		677
Quartz.........	Pure, crystallized	California	0.878		1.68	2.65	
Iron............	Du Pont	Pennsylvania	0.89		2.73		251

* Standard ball-mill tests made at other mesh sizes.
† Standard rod-mill tests made on this ore.

TABLE 1—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT 200-MESH—(*Continued*)							
			−200M.				
Cement rock	Cementos Quadalajara	Jalisco, Mexico	.892		1.305		1618-C
Gold	Little Long Lac	Ontario	0.903*	1	1.69	2.64	570
Gold	Francoeur Gold Mines Ltd.	Arntfield, Quebec	.91		1.72		1627
Sand	Krebs Pigment	Maryland	0.92		2.79		392
Cement clinker	N. A. Portland Cement Corp.	Security, Md.	.924		1.68		1114-A
Copper	New Cornelia Ore	Ajo, Ariz.	0.942		1.63	2.68	684
Gold	Galigher Co.	Utah	0.943		1.65		889
Cement raw mat	Cementos del Valle	Cali, Colombia	.95		1.51		1670
Gold	Powell-Rouyn	Quebec	0.960*		1.78		949
Gold	Santa Maria del Oro	Mexico	0.981		2.08	3.10	574
Taconite tails	Reserve Mining Co.	Babbitt, Minn	.98		2.15		1456
Copper	Anaconda	Montana	0.990*		2.18	3.23	910
Gold	M. A. Smith	Cuba	1.025		1.85		710
Gold	Malartic Goldfields Ltd.	Halet, Quebec	1.03		1.74		1592
Gold	Portland	Colorado	1.035		1.63	2.69	
Copper	Castle Dome	Miami, Ariz.	1.036*		1.69		1042
Cement rock	Cia. de Cementos Portland Diamante	Colombia	1.04		1.79		1642
Silver	Cia. Min. Carlota	Chile	1.042		1.88	3.00	660
Gold	Berens River Mines	Manitoba	1.045		1.95		705
Fe-Si, 25–75	Electro-Met. Corp.	New York	1.075*		1.94		1067
Phosphate rock	Am. Cyamid Co.	Brewster, Fla.	1.07		1.66		1221-C
Ti	Vanadium Corp. Am.	Pennsylvania	1.08		2.44		281
Ferro-silicon	Giffels & Vallet	Detroit, Mich.	1.075		1.94		1069
Copper	Morenci	Arizona	1.085*†		1.57	2.63	913
Cement clinker	Calaveros Cement Co.	Cal.	1.09		1.61		1590
Gold	Upper Can. Gold Mines	Ontario	1.097		1.88		821
Manganese	Gen. Manganese Corp.	S. Dakota	1.136		1.62		734
Copper	Miami Ore	Miami, Ariz.	1.139		1.64	2.69	
Lead-zinc	St. Joseph Lead Co.	Emerville, N. Y.	1.14		2.10		1657
Gold	Rochester-Plymouth	Nevada	1.141*		1.83		518
Limestone	N. A. Cement Co.	Security, Md.	1.15		1.8		1114-A
Copper	Silver Bell	Arizona	1.152*		1.68		1050
Gold	Kerr-Addison	Ontario	1.153*		1.83		799
Gold	Baguio Gold Min.	P.I.	1.160		1.62	2.63	402
Copper	Anaconda Copper Co.	Anaconda, Mont.	1.175		1.807		1477-A
Iron	U. of Minn., Oglebay, Norton & Co.	Babbitt, Minn.	1.17		2.40		1167-A
Gold	Kerr-Addison	Ontario	1.175*		1.81		799
Quartz	White Quartz—Sulphides	California	1.20		1.70		725
Taconite heads	Reserve Mining Co.	Babbitt, Minn.	1.20		2.15		1156
Tin	Prof. Gaudin—M.I.T.	Cambridge, Mass.	1.21		2.30		1254
Gold	Cline Lake	Ontario	1.225*		1.78		745
Iron	Moose Mountain	Ontario	1.227*		2.44	3.42	756
Arsenic-gold	Getchell Mines, Inc.	Red House, Nevada	1.23		1.70		1241
Copper	Utah Copper—Arthur	Utah	1.23*†		1.72		938
Copper	Nevada Consolidated Copper Co.	Chino Mines, Hurley, N. M.	1.24		1.83		1060
Feldspar	Golding-Keene Co.	Keene, N. H.	1.25		1.62		1553
Limestone	H. J. Kaiser Co.	California	1.257*		1.66		877
Gold	Homestake	S. Dakota	1.26*		1.91	3.12	
Copper	Cyprus Mines Corp.	Cyprus	1.284*		2.69		432
Coral limestone	E. P. Halliburton	Caracas, Venezuela	1.29		1.47		1605
Copper	Shale—White Pine	Michigan	1.304*		1.79	2.97	1000
Zinc	New Jersey Zinc—Conc.	Penna.	1.315		2.51		849
Taconite	Reserve Mining Co.	Babbitt, Minn.	1.32		2.15		1456
Gold	So. Amer. Dev. Co.	Ecuador	1.323*		2.67	3.81	752
Cement raw mat	H. W. A.—HSIN Cement Co.	Shanghai, China	1.37		2.71		1715
Cement raw mat	Mo. Portland Cement Co.	St. Louis, Mo.	1.39		1.78		1318
Lead-zinc	Axerio-Monteponi	Italy	1.396		2.88	4.43	562
Copper	Southwestern Eng. Co.	Los Angeles, Calif.	1.417		4.45		1717
Gold	C. M. & S. Co.—Golden Rose	Ontario	1.42*		2.10		864

* Standard ball-mill tests made at other mesh sizes.
† Standard rod-mill tests made on this ore.

TABLE 1—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TEST AT 200-MESH—(*Continued*)							
			−200M.				
Zinc	New Jersey Zinc Co.	Penna.	1.482		2.55		503
Clay	H. J. Kaiser Co.	California	1.585*		1.36		877
Gold	E. Malartic	Quebec	1.601*		1.87	2.80	779
Magnesite "NEG"	Magnesium do Brazil Ltda.	Rio de Janeiro	1.62		2.68		1698
Limestone	Calaveras Cement Co.	Cal.	1.66		1.64		1590
Limestone	Lawrence Cement	Pennsylvania	1.667*		1.90		972
Magnesium	Basic Refr. Inc.	Ohio	1.682*		2.13	2.93	1033
Gold	Preston East Dome	Ontario	1.69*		1.78		604
Cement raw mat.	Universal Atlas Cement Co.	Northampton, Pa.	1.71		1.84		1545
Magnesite "ORO"	Magnesium do Brazil Ltda.	Rio de Janeiro	1.81		1.665		1698
Gold	Butte Highlands	Montana	1.81*		1.66		861
Copper	Bingham Canyon	Utah	1.854		2.62	4.65	
Copper	Mines De Bor	Yugoslavia	1.94		2.1		249
Cement raw mat.	Cemento del Caribe	Barranguilla, Colombia	1.94		2.1		1529
Cement raw mat.	Mo. Portland Cement Co.	St. Louis, Mo.	1.944		1.87		1345
Manganese	L. G. Aguilar & Co.	Cuba	1.944		1.87		1028
Cement raw mat.	Mo. Portland Cement Co.	St. Louis, Mo.	2.06		1.78		1656
Cement raw mat.	Iraq Cement Co.	Bagdad, Iraq	3.51		1.69		1652
Ferro phosphorus	Monsanto Chemical Co.	Columbia, Tenn.	3.62		4.25		1262
Lead slag	Arcade Sm. & Ref.	Mass.	5.16		4.53		360
Avg.			1.00				

* Standard ball-mill tests made at other mesh sizes.

TABLE 2—*Standard Rod-mill Grindability Tests*

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
Tests at Minus 3-mesh							
			−3 M.				
Gravel	Arizona Sand & Stone Co.	Phoenix, Ariz.	7.26	7.7		2.65	1767
Clinker	Volunteer Cement Co.	Tennessee	29.34*	2.82	1.79		828
Granite	W. A. Burton	Texas	31.6	2.76	1.60	2.65	875
Avg.			22.73	4.4			
Tests at Minus 4-mesh							
			−4 M.				
Gravel	Warner Co.—Van Sciver	Pennsylvania	22.2	5.71	1.69		815
Iron	Charleston Iron Min.	Minnesota	25.2	9.2	2.34		754
Clinker	Volunteer Cement Co.	Tennessee	26.15*	2.67	1.79		828
Bauxite	Republic Min. & Mfg.	Arkansas	37.0	17.50	1.51		1053
Avg.			27.63	8.8			
Tests at Minus 6-mesh							
			−6 M.				
Gold	Lake Shore Gold Mines Ltd.	Kirkland Lake, Ontario	13.1	9.9	2.66		1644-B
Gravel	Material Service Corp.	Illinois	22.2	8.55	1.60		765
Iron	Warren Pipe & Foundry	New Jersey	49.5	5.72	2.08		1065
Calcite	New England Lime	Mass.	133.6	4.02	1.77		1066
Dolomite	New England Lime	Connecticut	319.0	2.39	1.85		1066
Avg.			107.48	6.1			
Tests at Minus 8-mesh							
			−8 M.				
Fused alumina	Monsanto Chemical Co.	St. Louis Mo.	7.37	10.3	2.66		1776
Gold	Lake Shore Gold Mines Ltd.	Kirkland Lake, Ontario	11.28	11.0			1644-B
Silicon-carbide	Exolon Co.	New York	37.10*	1.92	1.60	3.17	1052
Phosphate	Federal Chem. Co.	Tennessee	41.50*	24.36	1.80		814
Coal Slag	H. B. Reed, Inc.	Indiana	208.5	0.48	1.63		1026
Avg.			9.61	61 1			
Tests at Minus 10-mesh							
			−10 M.				
Limestone	Crushed Rock Prod. Co.	New York	1.63	9.25	1.67		906
Brick	Cohart Refr. Inc.	Kentucky	2.65*	16.73	2.22		809
Shale	Minn. Mining & Mfg. Co.	Wausau, Wisconsin	5.33	16.5	1.52		1217
Gold	Lake Shore Gold Mine Ltd.	Kirkland Lake, Ontario	9.3	12.0	2.66		1644
Tile scrap	Trenton Pottery Co.	Trenton, N. J.	9.68	.37	1.24		1314-A
Gold	Malartic Goldfields Ltd.	Halet, Quebec	9.7	18.9	1.74		1952
Coke	Kaiser Co. (Iron Steel Div.)	Fontana Calif.	9.72	16.5	.57		1043-A
Limestone	Pitts. Limestone Corp.	Pennsylvania	9.85	13.0	1.50		723
Copper ore	Cons. Mining & Smelting Co. of Canada	Kimberley, B. C.	10.8	23.8	2.29		1692
Gravel	Dravo Corp.	Pennsylvania	11.64	12.98	1.64		750
Limestone	Franklin Limestone Co.	Tennessee	12.65	15.20	1.65		953
Clinker	Volunteer Cement Co.	Tennessee	13.03*	7.72	1.79		828
Glass	Corning Glass Wks.	New York	14.15	5.17	1.01		945
Copper sink conc.	Cons. Mining & Smelting Co. of Canada	Kimberley, B. C.	16.70	21.0	2.64		1692
Sod. silicate	Du Pont	Indiana	16.80	7.61	0.94		954
Sand	Quinn Robbms Co. Inc.	Boise, Idaho	19.13	5.1	1.50		1235-A
Gravel	Industrial Silica Corp.	Youngstown, Ohio	19.28	1.0	1.68		1204-A
Chrome	Tekirova Madenleri Co.	Turkey	37.40	11.74	2.05		843
Avg.			12.74	11.9			
Tests at Minus 14-mesh							
			−14 M.				
Gold	Buffalo Ankerite Gold Mines Ltd.	S. Porcupine, Ont.	5.58	18.8	1.64		1639
Copper	Granby Consolidated		5.62	21.1	1.45		1566
Trap rock	Great Notch Granules Co.	Great Notch, N. J.	6.52	12.7	1.57		1440

* Rod-mill tests made at other mesh sizes.

TABLE 2—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT MINUS 14-MESH—(*Continued*)							
			−14 M.				
Gold	Waite-Amulet Mines Inc.	Noranda, Quebec	6.83	18.3		3.35	1761
Gold	Lake Shore Gold Mines Ltd.	Kirkland Lake, Ont.	7.36	15.3	1.62		1644
Slate	Maine Development Comm.	Augusta, Me.	7.50	16.5	1.41		1679
Radium	Eldorado Gold Mines	Ontario	7.76	17.66	1.95	2.77	1063
Tile	Arketex Ceramic Corp.	Indiana	8.15	18.02	1.60		766
Gold	Kerr-Addison Mines Ltd.	Virginiatown, Ontario	8.82	22.0		2.77	1765
Taconite	Reserve Mining Co.	Montreal, Wis.	8.91	16.7		3.42	1748
Copper ore	Cons. Smelting & Mining Co. of Canada	Kimberley, B. C.	9.66	15.0	2.29		1692
Limestone	Warner Co.	Phila., Pa.	10.03	3.3	1.69		1483
Fluorspar	Kinetic Chem. Inc.	New Mexico	11.15*	14.6	1.97		763
Copper	Utah Copper Co.	Arthur, Utah	11.88	24.2	2.70		1763
Magnetite	National Lead	Tahawus, N. Y.	12.21	10.2	4.46		1757
Calcines	Basic Dolomite Co.		13.50		1.62		583
Copper	Kennecott Copper Co.	Ray, Arizona	14.7	25.8	2.65		1752
Feldspar	Golding Keene Co.	N. Hampshire	14.90*	13.85	1.78		784
Limestone	Limestone Product Corp. of America	Newton, New Jersey	16.35	10.9	2.78		1700
Copper sink conc.	Cons. Smelting & Mining Co. of Canada	Kimberley, B. C.	16.40	24.1	2.64		1692
Fontana rock	T.V.A. Fontana Dam	Knoxville, Tenn.	19.2	14.3	2.65		1750
Nickel matte	Int. Nickel Co.	W. Virginia	20.96	13.98	3.81		1009
Lead-zinc	St. Joseph Lead Co.	Emeryville, N. Y.	22.4	12.7	2.32		1657
Magnetite	Jones-Laughlin Ore Co.	Star Lake, N. Y.	28.5	7.9	1.99	3.23	1644
Iron	Jones-Laughlin Ore Co.	Star Lake, N. Y.	30.3	11.9	1.99		1644-B
Copper	Tennessee Copper Co. London Mill	Copperhill, Tenn.	31.25	19.9		4.52	1750
	Pickands-Mather Co.,						
Langbeinite	Union Potash & Chem.	New Mexico	40.90*	8.77	1.59		842
Slag	Celotex Corp.	Ohio	163.5		0.36		919
Avg.			20.02	15.71			
TESTS AT MINUS 20-MESH							
			−20 M.				
Graphite	U. S. Graphite Co.	Michigan	2.22		1.24		728
Brick	Cohart Ref. Co.	Kentucky	2.57*	20.61	2.22		809
Iron	Mozan	Japan	4.20†	20.19	2.41	3.42	914
Alumina	Exolon Co.	New York	5.19	10.84	2.65		1052
Silicon carbide	Exolon Co.	New York	7.50	6.73	1.60		1052
Slate	Maine Development Comm.	Augusta, Maine	7.50	16.5	2.37		1679
Gravel	Industrial Silica Corp.	Youngstown, Ohio	7.61	8.6	1.68		1204-A
Slag	Ohio Ferro-Alloys	Ohio	8.40	21.82	1.99		933
Sand	Walter C. Best Co.	Chardon, Ohio	8.73	8.5	1.73		1305
Quartzite	Smith & Koelliker	Ohio	9.26	15.43	1.84		774
Copper	Utah Copper—Arthur	Utah	9.90*†	25.07	1.72		938
Titanium	Nat. Lead Co.	Missouri	11.50*	13.0	2.64		1017
Rutile	American Rutile Corp.	Virginia	11.95	31.75	1.81		971
Feldspar	Golding Keene Co.	N. Hampshire	13.34*	16.30	1.78		784
Copper	Morenci Ore	Arizona	14.98*†	23.75	1.57	2.63	913
Feldspar	Cons. Feldspar Corp.	Tennessee	16.36	14.45	1.52		934
Iron	Rep. Steel—Harmony	New York	18.60†	7.16	2.19		822
Bauxite	Porocel Corp.	Arkansas	21.27	19.60	1.25		884
Iron	Rep. Steel—Old Bed	Mineville, N. Y.	25.95†	7.05	2.70		822
Iron	Rep. Steel—New Bed	Mineville, N. Y.	28.22†	7.05	2.71		822
Lime rejects	Louisville Cement Co.	Milltown Ind.	38.1	2.42	.715		1090
Barite	Barium Min. Corp.	W. Virginia	59.5	15.7	3.14		1007
Ferro-silicon	Southern Ferro Alloys Co.	Chattanooga, Tenn.	62.8	6.4	2.16		1574
Avg.			17.20	14.5			

* Rod-mill tests made at other mesh sizes.
† Standard ball-mill tests made on this ore.

TABLE 2—(*Continued*)

Ore	Description by	Location	Net Grams Undersize per Revolution	−200-mesh in Product, Per Cent	Specific Gravity		No.
					Apparent	True	
TESTS AT MINUS 28-MESH							
			−28 M.				
Cr Metal	Electro-Met. Co.	New York	1.90†	3.00	3.91	7.01	1008
Sod. silicate	Diamond Alkali Co.	Ohio	4.72	15.56	1.62		739
Glass	Pitts. Plate Glass Co.	Pennsylvania	4.97	10.47	1.49	2.60	746
Feldspar	Consolidated Feldspar Corp.	Erwin, Tenn.	7.39	15.8	1.65		1103-A
Manganese	E. W. Creevy	Costa Rica	7.45	24.00	2.02		968
Gypsum	Diamond Crystal Salt		7.77	41.60	1.90		696
Copper	Utah Copper—Arthur	Utah	7.93*†	31.21	1.72		938
Titanium	Nat. Lead Co.	Missouri	9.165*	17.12	2.64		1017
Iron	M. A. Hanna Co.—Clifton	New York	9.30†	15.00	2.76		1022
Copper	Morenci Ore	Arizona	11.98*†	28.86	1.57	2.63	913
Ferrosilicon	Pitts. Met. Co.	New York	20.92*	10.36	2.77		839
Avg.			8.50	19.3			
TESTS AT MINUS 35-MESH							
			−35M.				
Pea gravel	Am. Aggregates Corp.		3.70	29.8	1.68		1649
Iron	Mr. W. W. Christy	Malvern, Ark.	4.24	24.8	1.65		1104
Pumice	Barnsdall Tripoli Co.	Missouri	4.52	58.4	0.64		749
Zinc	New Jersey Zinc Co.	N. Jersey	5.73	26.0	2.39		930
Phosphate	Federal Chem. Co.	Tenn.	5.87	37.84	1.38		533
Copper	Morenci Ore	Arizona	6.45*†	39.64	1.57	2.63	913
Copper	Utah Copper—Arthur	Utah	6.62*†	35.53	1.72		938
Gold	Mineral Min. Corp.	S. Carolina	7.09†	54.61	1.43		782
Foam slag	Bethlehem Steel Corp.	Sparrows Point, Md.	7.70	30.9	.74		1119-A
Feldspar	Consolidated Feldspar Corp.	Erwin, Tenn.	8.33	22.5	1.60		1103
Limestone	Limestone Products Corp. of Amer.	Newton, N. J.	8.70	21.6	1.94		1395
Iron	Rep. Steel—Old Bed	New York	9.15*†	18.79	2.70		822
Limestone	Limestone Products Corp. of Amer.	Newton, N. J.	9.38	22.8	2 78		1700-B
Iron	Rep. Steel—Harmony	New York	9.43*	17.58	2.20		822
Potash ore	Mines Domainales de Potasse D' Alsace	Mulhous, France	10.1	22.1	2.15		1716
Sylvanite	Union Potash & Chem. Co.	New Mexico	11.22*	20.21	1.30		842
Ferrosilicon	Pitts. Met. Co.	New York	13.68*	24.26	2.68		839
Barite	United Pigment—Meggen	N. Jersey	21.01	37.09	2.98		848
Pitch	Crosset Chem. Co.	Arkansas	35.2	26.24	1.27		813
Barite	United Pigment—Tenn.	N. Jersey	73.43	39.94	3.09		848
Avg.			13.07	30.5			
TESTS AT MINUS 48-MESH							
			−48M.				
Copper	Amygdaloid	Michigan	1.995†	48.94	1.87	2.93	1000
Copper	Sandstone Vein	Michigan	3.733†	39.40	1.83	2.68	1000
Gold	Seal Harbor Gold Mines	Nova Scotia	4.66	40.4	1.78	2.76	602
Kyanite	Phosphate Recovery Corp.	Virginia	4.77	20.46	1.74	2.82	760
Copper	Morenci Ore	Arizona	4.89*†	44.87	1.57	2.63	913
Garnet ore	Barton Mines Corp.	N. Creek, N. Y.	5.25	20.5	1.95		1510
Fluorspar	Kinetic Chem. Inc.	New Mex.	5.40*	30.5	1.96		763
Sylvanite	Union Potash & Chem.	New Mex.	5.43	28.21	1.23		842
Copper	Utah Copper—Arthur	Utah	5.43*†	43.42	1.72		938
Langbeinite	Union Potash & Chem.	New Mex.	5.73*	31.27	1.59		842
Tripoli	Western Minerals, Inc.	Kansas	6.89†	77.35	1.14		883
Sinter	Aluminum Co. of Can.	Quebec	8.75*†	36.82	0.84		1080
Cryolite	Penna. Salt Mfg. Co.	Pennsylvania	13.93	37.62	1.92		958
Avg.			5.91	38.4			
TESTS AT MINUS 65-MESH							
			−65M.				
Magnetite	National Lead Co., MacIntyre	Tahawus, New York	2.47	43.0	2.65	4.46	1757
Fe-Si, 50–50	Electro-Met. Co.	New York	2.92	39.18	3.08		947
Molybdenum	Molybdenum Corp. of Amer.	Urad, Colo.	2.92	64.5	3.08		1140-A
Sinter	Aluminum Co. of Can.	Quebec	6.25*†	41.45	0.84		1080
Avg.			3.64	47.0			

* Rod-mill tests made at other mesh sizes.
† Standard ball-mill tests made on this ore.

TABLE 3—*Impact Crushing Tests*

Test No.	Name and Location	Material	Compressive Strength, Psi Avg	Specific Gravity	Ft-lb per Inch	
					High	Average
1672	Standard Oil Co., Esso Lab., Baton Rouge, La.	Black porous iron sinter			1.2	0.61
1716	Mine Domaniaies de Potasse, Mulhouse, Alsace, Fr.	Potash salts		2.15	4.4	2.84
1519	Lawrence Portland Cement Co., Thomaston, Me.	Cement clinker	12530	3.15	9.9	3.07
1675-A	Martin Bros., Butte, Mont.	Pyritic gold ore (light color)		2.5	5.6	3.48
1536	Pennsylvania Salt Mfg. Co., Natrona, Pa.	Siderite			8.6	3.48
1513	A. C. Bateman, Johannesburgh, S. A.	Limestone		2.6	7.7	3.55
1698	Chaves Y Cia, Magnesium do Brazil, Rio de Janeiro, Br.	Magnesite "ORO"			5.9	3.67
1536	Pennsylvania Salt Mfg. Co., Natrona, Pa.	Cryolite			5.9	3.82
1516	Saticoy Rock Co., Saticoy Cal.	Granite pebbles		2.6	7.3	4.73
1715	HWA-HSTN Cement Co., Shanghai, China	Argillaceous stratified limestone			6.3	4.82
1394	St. Claire Lime, Okla. City, Oklahoma	Limestone		2.6	8.2	5.00
1341	Portage Manly Sand Portage, Wis.	Sandstone		2.6	6.4	5.22
1536	Pennsylvania Salt Mfg. Co., Natrona, Pa.	Silica and fluorspar			9.6	5.27
1715	HWA-HSTN Cement Co., Shanghai, China	Pure limestone			7.9	5.48
1379	Wm. Knight, N. Carolina	Magnetite		4.77	5.6	4.77
1377	Republic Steel, Spaulding, Ala.	FeO (fine)		3.30	10.8	5.70
1675-A	Martin Bros., Box 76, Butte, Mont.	Pyritic gold ore, (dark color)	24550	2.7	9.8	6.02
1402	Petoskey P. C. C., Petoskey, Mich.	Fine limestone	14220	2.58	12.0	6.09
1407	Champion Spark Plug, Detroit, Mich.	Andalusite	10207	3.08	8.0	6.18
1324	Mississippi Lime Co., St. Genevieve, Mo.	White limestone	10285	2.6	7.8	6.28
1484	Southwest Stone Co., Chico, Texas	White limestone		2.68	10.3	6.31
1750	Tennessee Copper Co., Copperhill, Tenn.	London mill ore		4.52	8.8	6.43
1377	Republic Steel, Spaulding, Ala.	Iron ore		3.29	10.8	6.58
1416	LaClede Christy, St. Louis, Mo.	Calcined kyanite ore		2.89	9.8	6.62
1661	Glen Falls Portland Cement Co., Glen Falls, N. Y.	Limestone	12546	2.70	12.00	6.84
1720	Ververies d'Egypt, Cairo, Egypt	Quartz	7620	2.88	16.00	6.92
1516	John D. Gregg, Roscoe, Cal.	Granite pebbles		2.6	11.8	7.07
1345	Mo. Portland Cement Co., St. Louis, Mo.	Limestone	16100	2.6	12.1	7.18
1613	DuPont, Terre Haute, Ind.	Pyrite in coal		3.6	16.0	7.55
1736	Fabrica Nacional de Yidrio, Caracas, Venezuela	White quartz stones		2.67	13.2	7.85
1567	Cedar Bluff Stone Co., Princeton, Ky.	Soft Ag. Limestone		2.6	11.3	7.97
1516	Graham Bros., Inc., El Monte, Cal.	Granite pebbles		2.6	19.3	8.37
1384	Loomis Talc Co., Coveneua, N. Y.	Talc		2.83	11.4	8.55
1709	Major Wm. Knight, Lakeland, Fla.	Limestone rock	11525		13.8	8.94
1388	Jones & Laughlin, Star Lake, N. Y.	Iron ore			15.5	9.84
1334	Southern Ferro, Chattanooga, Tenn.	Ferrosilicon		6	14.8	10.16
1366	Wisconsin Steel, Nashwauk, Minn.	Hard ore		4.20	21.7	10.16
1750	Tennessee Copper Co., Copperhill, Tenn.	Isabella mill ore		4.52	16.0	10.36
1497	Southwest Stone Co., Springtown, Okla.	Limestone		2.6	18.8	10.47
1533	General Crushed Stone Co., Auburn Plant	Limestone		2.6	19.5	10.74
1480	Southwest Stone Co., Knippa, Texas	Black trap rock		3.12	16.0	10.90
1406	Arizona Sand & Rock, Phoenix, Arizona	Pebbles	28835	2.6	17.7	11.14
1611	Superior Stone Co., Red Hill, Va.	Granite	15020	2.82	13.4	11.20
1347	Weston & Brooker, Columbia, S. Carolina	Granite (Cayce)	17923	2.6	14.8	11.42
1358	Union Steel Castings, Pittsburgh, Pa.	Fe-Mn-C alloy		7.21	15.5	11.55
1567	Cedar Bluff Stone Co., Princeton, Ky.	Limestone Ag.		2.6	15.5	11.60
1398	Icaza & Co., Panama	Limestone		2.6	17.3	11.66
1719	Weston & Brooker Co., Columbia, S. Carolina	Red granite (Cayce)	26430	2.63	15.1	11.79
1689	U. S. Engineers Office, Jacksonville, Fla.	Granite	16475	2.63	16.4	11.80
1719	Weston & Brooker Co., Columbia, S. Carolina	Light gray granite (Cayce)	26070	2.61	15.2	11.81
1567	Cedar Bluff Stone Co., Princeton, Ky.	Limestone		2.6	16.0	12.07
1552	A. S. & R. Co., Tacoma, Wash.	Limestone	16685	2.74	17.5	12.20
1750	Tennessee Copper Co., Copperhill, Tenn.	Fontana rock		2.65	21.9	12.3
1763	Utah Copper Co., Arthur, Utah	Copper ore		2.7	19.8	12.58
1487	Great Western Sugar Co., Horse Creek, Wyo.	Limestone	27750	2.6	20.5	12.60
1772	Granite Rock Co., Watsonville, Calif.	Granite	9330		15.5	12.70
1367	Batesville, White, Ark.	Limestone		2.6	22.2	13.08
1324	Mississippi Lime Co., St. Genevieve, Mo.	Gray limestone	14410	2.6	19.7	13.16
1412	Cold Springs Granite Co., Minn.	Pink granite		2.6	13.6	13.48

TABLE 3—(*Continued*)

Test No.	Name and Location	Material	Compressive Strength, Psi Avg	Specific Gravity	Ft-lb per Inch	
					High	Average
1536	Pennsylvania Salt Mfg. Co., Natrona, Pa.	Granite		2.6	18.0	13.65
1761	Waite-Amulet Mines Inc., Noranda, Quebec, Canada	Gold ore		3.35	24.3	13.86
1719	Weston & Brooker Co., Columbia, S. Carolina	Gray granite (Cayce)	23100	2.65	15.6	13.97
1560	Climax Molybdenum Co., Climax, Colo.	Molybdenum ore	19970	2.62	19.4	14 09
1427	Oliver Iron Mining, Tower, Minn.	Jasper		3.42	25 2	14.51
1412	Concrete Materials, Sioux Falls, S. D.	Granite		2.6	18.7	14.54
1469	Helena Sand Gravel Co., Helena, Mont.	Trap rock		2.81	24.0	14.69
1402	Petoskey P. C. C., Petoskey, Mich.	Coarse limestone	16785	2.63	27.0	14.75
1398	Icaza & Co. Panama	Sandstone		2.6	22.1	14.88
1396	Correale Const. Co., Minnersville, Pa.	Shale	8106	2.66	22.1	15.06
1719	Weston & Brooker Co., Camak, Ga.	Trap rock (Camak)	17650	2.79	20.0	15.27
1719	Weston & Brooker Co., Camak, Ga.	Granite (Cayce)	25840	2.64	34.7	15.28
1621	General Crushed Stone Co., Quakertown, Pa.	Trap rock	15967	3.04	25.2	15.35
1567	Cedar Bluff Stone Co., Princeton, Ky.	Limestone		2.6	19.0	15.35
1719	Weston & Brooker Co., Columbia, S. Carolina	Red granite (Cayce)	25270	2.64	20.5	15.62
1397	Hanna Ore Co., De Grasse, N. Y.	Iron ore	24305	4.24	23.4	15.64
1372	T.V.A. Fontana Dam, Tenn.	Limestone		2.6	30.5	15.06
1719	Westone & Brooker Co., Columbia, S. Carolina	Gray coarse granite (Cayce)	28030	2.66	20.5	16.02
1565	John T. Dyer (A-C office), Harrisburg, Pa.	Gray granite		2.98	25.3	16.08
1502	Lynn Sand & Stone Co., Mass.	Rhyolite olivine		2.65	21.1	16.11
1390	Old Colony Crushed Stone, Quincy, Mass.	Granite		2.6	28.7	16.59
1578	California Rock & Gravel, Calif.	Trap rock gravel	17900	2.8	28.8	16.74
1571	Globe Iron Co., Duluth, Minn.	Specular hematite	33070	2.9	29.8	17.03
1710	Palmetto Quarries, Columbia, S. Carolina	Granite	20180		23.2	17.80
1318	Missouri Portland, Batesville, Ark.	Limestone	20345	2.6	33.2	17.91
1318	Missouri Portland, St. Louis, Mo.	Limestone	19222	2.6	23.8	18.00
1803	Pickands Mather Co., Hibbing, Minn.	Iron(Pit 5D)	23784	3.36	27.1	18.39
1748	Reserve Mining Co., Montreal, Wis.	Babbitt taconite		3.42	25.6	18.81
1505	Koppers Co., Kobuta, Pa.	Al-Ni pigs		3.70	27.5	19.15
1456	Reserve Mining, Babbitt, Minn.	Taconite	41360	3.16	36.6	19.88
1502	Lynn Sand & Stone Co., Boston, Mass.	Gabbro diorite		2.85	29.8	20.07
1363	Great Notch Granule Co., N. J.	Trap rock			28.9	20.72
1412	Cold Springs Granite Co., Morton, Minn.	Red granite		2.6	27.1	21.85
1412	Cold Springs Granite Co., Morton, Minn.	Granite		2.6	27.1	21.90
1412	L. G. Everist Co., Del Rapids, S. D.	Everist granite		2.6	28.5	22.20
1719	Weston & Brooker Co., Columbia, S. Carolina	Dark gray granite (Cayce)	26170	2.65	34.7	22.38
1708	Southwestern Stone Co., Knippa, Texas	Trap rock	29287		27.0	23.9
1757	National Lead Co., Tahawus, N. Y.	Magnetite		4.46	30.8	24 30
1612-A	Basic Ind. Lab. (A-C), Milwaukee, Wis.	Ferrosilicon (15 % Si)		6.59	31.6	25.02
1358	Union Steel Castings, Pittsburgh, Pa.	Fe-Mn alloy		6.84	34.1	25 20
1515	Oliver Iron Mining Co., Soudan, Tower, Minn.	Hematite	38550	4.85	34.0	25.29
1358	Union Steel Casting Co., Pittsburgh	Si-Mn alloy		6.63	29.6	29.60
1803	Pickands Mather Co., Hibbing, Minn.	Iron (Pit 11E)	40975	2.92	58.7	30.34
1427	Oliver Iron Mining Co., Soudan, Tower, Minn.	Hematite		4.93	40.8	31.32
1412	Spencer Quarries Co., S. Dakota	Red granite		2.6	33.9	32.00
1407	Champion Spark Plug, Detroit, Mich.	Dumortierite	42233	3.21	60.2	38.10
1493	Vanadium Corp. of America, Niagara Falls, N. Y.	Chrome metal		7.36	73.8	55.45
	Average		21381		20.07	13.53

TABLE 4—*Comparative Open-circuit Grindability Tests and Standard Grindability Tests 200-mesh, October* 23, 1946

Open-circuit Tests

First Stage (I) Large Ball Mill			Second Stage (II) Trojan Jar Mill		
Mesh	Limits	Rev.	Mesh	Limits	Rev.
10	25–95		100	60–99	
20	20–85		200	50–92	
	Average			Average	

Standard Closed-circuit Grindability Tests Made at 200-mesh

Test No.	Name and Location	Wet or Dry Grind	Material	Grindability, 200-mesh	Index I[a]	Index II[a]
1652	Iraq Cement Co., Bagdad, Iraq	Wet	P. C. mix	3.51	878.0	2812.0
1316	Cementa del Caribe, Barranquilla, Colombia	Wet	Limestone		1049.0	2938.0
1448	Cemento del Caribe, Barranquilla, Colombia	Wet	Limestone		862.0	3187.0
1642	Cia de Cemente Portland Diamente, Colombia	Wet	Limestone	1.04	2026.2	3245.0
1492	Fabrica del Cemento Samper, Bogota, Colombia	Wet	P. C. mix		1612.5	3312.0
1715	HWA-HSIN Cement, Shanghai, China	Dry	P. C. mix	1.37	1900.0	3397.0
1600	La Cruz Azul, Jasso Hgo, Mexico	Wet	P. C. mix		1732.0	3545.0
1590	Calavaras Cement Co., San Francisco, Calif.	Wet	Limestone	1.66	1245.0	3625.0
1114-A	North American Cement Corp., Alsen, N. M.	Wet	Limestone		1631.2	3750.0
1585	Fabrica del Cemento Samper, Bogata, Colombia	Wet	P. C. mix		4850.0	3875.0
1656	Missouri Portland, St. Louis, Missouri	Wet	P. C. mix	2.06	1048.0	3916.0
1605	Halliburton, Caracas, Venezuela	Wet	P. C. mix	1.29	1637.0	4002.0
1518	Cemento, Panama, Cristobal, Panama	Wet	P. C. mix		470.0	4050.0
1114-A	North American Cement Corp., Alsen, N. Y.	Wet	Limestone		1737.5	4062.5
11-34	Universal Atlas-Leeds, Alabama	Wet	P. C. mix		1642.0	4130.0
1715	HWA-HSIN Cement Co., Shanghai, China	Wet	P. C. mix		1900.0	4187.5
1545	Universal Atlas—Northampton, Penn.	Wet	P. C. mix	1.71	1287.0	4218.0
1529	Cemento del Caribe, Barranquilla, Colombia	Wet	P. C. mix	2.47	819.0	4371.0
1529	Cemento del Caribe, Barranquilla, Colombia	Wet	P. C. mix	2.37	787.0	4413.0
1114-A	North American Cement Corp., Alsen, N. Y.	Wet	Limestone		2550.0	4487.5
1345	Missouri Portland, Prospect Hill, St. Louis	Wet	Limestone	2.68	1200.0	4625.0
1180-A	Union Portland Cement, Devils Slide, Utah	Wet	Limestone		2281.2	4625.0
1318	Missouri Portland, Memphis, Tenn.	Wet	P. C. mix	1.39	1042.0	4637.0
1114-A	North American Cement Corp, Alsen, N. Y.	Wet	Limestone		2068.5	5187.5
1189-A	Kelly Island Lime & Transport Co., Ohio	Wet	Limestone II		450.0	5825.0
1190	Lone Star, Nazareth, Penn.	Wet	Limestone		1975.0	6500.0
1189-A	Kelly Island Lime & Transport Co., Ohio	Wet	Limestone I		475.0	7062.5
1134	Universal Atlas—Leeds, Alabama	Dry	PCclinker		1979.0	7625.0
1670	Cementos Del Valle, Cali, Colombia	Wet	P. C. mix	0.95	1888.5	7662.5
1618-B	Cementos Guadalajara, Jalisco, Mexico	Dry	PCclinker		1942.9	7875.0
1698	Magnesium Do Brazil Ltda., Rio De Janiero	Dry	"ORO"	1.74	737.5	9375.0
1698	Magnesium Do Brazil Ltda., Rio De Janiero	Dry	"NEG"	1.62	807.7	9875.0
1795	Carlos T. Novella & Cia., Managua, Nicaragua	Dry	P. C. mix		1545.0	10375.0
11-34	Universal Atlas—Leeds, Alabama	Dry	Clinker		1703.0	14250.0
1267	Penn.—Dixie, Richard City, Tenn.	Dry	Clinker I		1435.0	16375.0
1618-A	Cementos Guadalajara, Jalisco, Mexico	Dry	P. C. mix		1900.0	16750.0
1222-A	Louisville Cement Co., Speed, Indiana	Dry	Clinker		2143.7	17500.0
1267	Penn.—Dixie, Richard City, Tenn.	Dry	Clinker II		1749.0	19000.0
1609	Santa Cruz P. C. Co., San Francisco, Calif.	Dry	Clinker		1692.2	20000.0
1114-A	North American Cement Corp., Hagerstown, Md.	Dry	Clinker II		1575.0	20500.0
1636	Riverside Cement Co., Oro Grande, Calif.	Dry	P. C. mix	1 56	1512.5	20625.0
1519	Lawrence Portland Cement, Thomaston, Me.	Dry	Clinker		1887.5	24125.0
1590	Calavaras Cement Co., San Francisco, Calif.	Dry	Clinker		1456.0	24500.0
1618-B	Cementos Guadalujara, Jalisco, Mexico	Dry	Clinker		988.7	24750.0
1114-A	North American Cement Corp., Hagerstown, Md.	Dry	Clinker I	0.924	2087.5	27875.0
1180-A	Union Portland Cement, Devil's Slide, Utah	Dry	Limestone		2950.0	28875.0
1617	Consolidated Portland Cement, Cement City, Mich.	Dry	Clinker	0.87	1126.0	31312.5
1795	Carlos T. Novella & Cia, Managua, Nicaragua	Dry	PCclinker		1656.0	37875.0
1114-A	North American Cement Corp., Alsen, N. Y.	Dry	Limestone	1.15	3437.5	44125.0

Impact and Compression

		Impact, Ft-lb per In.		Compression, Psi	
		Average	Maximum	Average	Maximum
1715	HWA-HSIN Cement Co., Shanghai, China	4.82	6.3		
1345	Missouri Portland Cement Co., Prospect Hill	7.18	12.1	16,100	18,120
1318	Missouri Portland, Memphis, Tenn.	18.0	23.8	20,545	24,480
1698	Magnesium Do Brazil Ltda, Rio De Janiero	3.67	5.9		
1519	Lawrence Portland Cement Co., Thomaston, Me.	3.07	9.88		

[a] Indices I and II equal the average revolutions required between the limits shown.

Wear Tests on Grinding Balls

By T. E. Norman,* Junior Member, and C. M. Loeb, Jr., Member AIME

(New York Meeting February 1948)

The use of ball, rod and tube mills for grinding ore, cement and other materials has grown so rapidly during the past forty years that the world's annual consumption of ferrous grinding media for these mills is now estimated to be between one half million and one million tons per year. Ferrous grinding balls constitute the major portion of this tonnage. Obviously they represent sufficient value to justify thorough studies of the factors governing their performance.

The selection of grinding balls is governed principally by: 1. Quality (wear resistance, impact resistance, soundness, and the like). 2. Sources of supply and delivered cost. 3. Grinding characteristics or efficiency in the ball mill. This paper deals principally with the quality of ferrous grinding balls. In the study of these factors certain data relative to the fundamental nature of ball wear in ball mills have been obtained. These data are also presented and discussed briefly.

The Development of a Suitable Wear Test

A study of the fundamental factors governing the quality of grinding balls has been hampered seriously by the fact that a competent test has, in the past, involved the purchase of several hundred tons of balls of a specific type which were then run in one or more ball mills for a period ranging from several months' to several years' duration. Often, during the period of test, it became necessary to change operating conditions or the character of the ore fed to the test mill with the result that the rate of ball consumption changed and the test figures became of little value. Under such circumstances, progress in the development of better grinding balls, has been necessarily slow.

Economic factors and variations in the quality of balls produced by different sources of supply generally make it necessary for each mill operator to determine for himself the most suitable type of balls for his mills. In our own ball mill grinding operations at Climax, Colo., we were faced with this problem. After we had run a few large scale wear tests at considerable expense we decided to investigate the possibilities of a small scale wear test which would be capable of testing numerous types of balls within a relatively short time.

The most important requirement of any test is, of course, that it give results which can be used to predict accurately the wear in full scale operations. It was known that Ellis and his associates[1,2] had developed a method of testing grinding balls by small scale tests run at the Ontario Research Foundation. Ellis' method of testing was used as a starting point in our investigations. In the course of our tests a number of modifications of the original method were found to be desirable so that, by an evolutionary procedure, a method of small scale

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[1] References are at the end of the paper.

testing has been developed which we believe makes possible an accurate estimation of the wear in full scale operations.

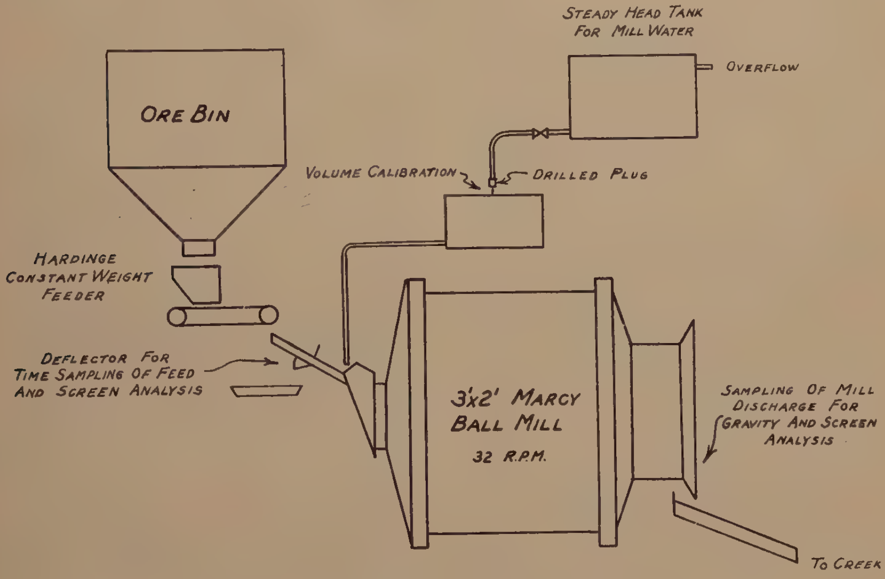

Fig 1—Flowsheet of wear tests at Golden.

Our first small scale tests were run at the Colorado School of Mines State Experimental Plant in Golden, Colo. These were run in a Marcy ball mill approximately 3 ft id by 2 ft long, lined with ship-lap steel liners and equipped with a discharge grate which could be adjusted by means of diaphragm rings to discharge the ground pulp at various levels. The discharge could also be sealed for batch grinding tests. Fig 1 illustrates the arrangement used for the test when operating on open circuit grinding with a continuous feed and discharge. This arrangement was used for most of our preliminary tests though in a few cases it was found desirable to run a series of batch tests for the study of certain variables. Abrasives used in the tests were crushed Climax ore, a river sand very similar in abrasive characteristics to Climax ore, a commercially pure crushed feldspar and a relatively pure type of crushed calcite.

The balls tested were, in most cases, nominally 3 in. in diam.

Our preliminary tests differed from Ellis' method principally in the matter of dimensions. We used balls 3 in. in diam in a mill 3 ft in diam where Ellis, in most of his tests, used balls 1 in. in diam in a mill 1 ft in diam or smaller.

Our tests in the 3-ft mill were found to be useful for preliminary testing or for the study of certain fundamental factors affecting ball wear. Usually the order of merit of a series of balls could be quite well established in such tests. The test was found, however, to have certain limitations. For instance, the impact conditions in the 3 ft mill did not duplicate those in our large 9 ft diam mills at Climax. Also, the spread in relative wear resistance between a good and a poor type of ball was generally different from that obtained in our larger mills at Climax, or in the mills at other mining operations.

Ellis[1] has demonstrated that the character of the abrasive has a marked influence on the relative wear resistance of various types of balls. This was confirmed in our preliminary tests. Ordinarily the hard abrasives such as quartz give a relatively small spread while the softer abrasives, such as feldspar and calcite, tend to produce a relatively large spread between the wear resistance of a good and a poor type of ball.

Because of the foregoing limitations of the tests in the 3 ft mill, a new testing technique was developed whereby many types of balls could be tested on a small scale in a commercial mill without seriously interfering with its regular operation. Results of such tests, when comparisons have been available, show excellent agreement with large scale tests made in the same mills where only one composition was used for the entire ball charge.

Detailed Procedure for Running Wear Tests in Commercial Mills

The technique used in running our wear tests, in its final stage of evolution, was as follows:

1. A series of groups of balls was selected with the balls in each group representing steel or iron of a specific type and treatment. A "group" usually consisted of from 5 to 15 balls. Our tests have indicated that if all the balls in any group are similar each ball will show, within the limits of experimental error, exactly the same weight loss per unit of area. There was very little advantage to be gained therefore from the use of large groups. One group of balls in each series was of the type used as a standard for comparison.

2. The balls in each group were marked with a distinctive mark, such as one or two notches or drilled holes or a combination of a notch and a drilled hole. Where two notches or drilled holes and notches were used on a ball, they were placed at definite angular distances from each other on the ball surface. Generally the marks were about ¼ in. deep with the holes ¼ in. diam and the notches $\frac{1}{16}$ or ⅛ in. wide by 1 in. long. The notches were cut with a small abrasive cut-off wheel.

Comparative tests have indicated that the one or two notches or holes placed in these test balls produced no measurable difference in rate of wear except in cases where spalling occurred at the edge of the notches. Where spalling did occur, it was generally of such a nature that the weight loss due to this spalling could be estimated and the necessary corrections made to determine the weight loss caused by wear alone.

3. After marking, the surface defects such as scaling and decarburization were removed from the test balls by a "wear-in" in a small ball mill for a sufficient time to remove metal to a depth of at least 0.040 in. below the original surface of the ball.

4. Where the groups of test balls were to be run in a large mill along with the regular charge from which it would be difficult to recover all the test balls, it was found necessary to adjust the weight (after the wear-in) of each ball in a group to an identical value. This was done by grinding on an abrasive grinding wheel. By having all balls in a group of equal weight at the beginning of the wear test it was unnecessary to recover all the balls in the group from the mill when the test was complete.

5. All the balls were weighed carefully both in air and while suspended in water containing a wetting agent. From these data the density, volume and surface area of each ball were calculated by assuming that the ratio of volume to surface area was equal to that of a perfect sphere.

6. The mill in which the wear test was to be run was selected and the entire series of test balls charged at once. They were allowed to run in this mill, along with the regular charge of balls under normal operating conditions, for a sufficient period of time to establish a reliable wear factor. In commercial mills the test balls were generally run for a sufficient length of time to wear off a layer of metal about ⅛ in. thick

from their surface. The mill was then stopped and the marked test balls were sorted out from the rest of the charge. Sorting was accomplished by dumping the entire charge of balls on the floor, or preferably in large mills, by having two or three men pick the marked balls from the surface of the ball charge inside the mill while it was slowly rotated 180° with a crane and rope.

The time for shutdown and recovery of balls from a test mill was generally chosen so that it would coincide with the time the mill was to be shut down for relining or other repairs. By doing this the wear test did not interfere with normal operations of the ball mill. We did not attempt to find 100 pct of the test balls. Generally, however, from 60 to 80 pct of them were found without difficulty within about 2 hr. During the search period a close watch was kept for test balls which had broken or spalled during the test.

7. The test balls were sorted, cleaned and weighed and their weight loss per unit of surface area was calculated. This figure was compared to the loss per unit of area on the standard balls included in the test. From this comparison an "abrasion factor" or relative rate of wear was calculated. For instance on a typical test the standard balls lost 116.0 g per 100 sq cm of original surface area. The balls in "group 1" showed an average loss of 128.7 g per 100 sq cm. The standard ball was always nominally assigned an abrasion factor of 100. The abrasion factor (relative rate of wear) of the balls in group 1 was, therefore, $\frac{128.7}{116.0} \times 100 = 111.0$. In this paper all abrasion factors are given to the nearest whole number.

All abrasion factors listed in this paper have been obtained by the foregoing procedure. In all cases our standard for comparison was a group of martensitic forged steel balls containing 0.75 to 0.88 pct carbon and 0.20 to 0.30 pct molybdenum. These standard balls had been made in regular commercial practice under carefully controlled conditions and were found to be very uniform in quality.

In studying the abrasion factors given in this review it should be clearly realized that they represent relative rates of wear. Balls with an abrasion factor higher than 100 wear away faster and are, therefore, poorer than the standard, while balls with an abrasion factor of less than 100 wear more slowly and are, therefore, superior to the standard.

The abrasion factor represents the relative rate of wear of that portion of the ball which was worn away during the test. If the ball is homogeneous from surface to center the abrasion factor should be representative of relative wear resistance of the entire ball. In some cases, however, the ball may be less wear resistant at its center than at its surface. When this condition exists, allowances must be made in evaluating the relative wear resistance of the entire ball. Generally, however, the correction necessary is very slight. For instance, if a steel ball 3 in. in diam is fully hardened to a depth of 1 in. below its surface, then the weight of hardened steel in the ball is 96.3 pct of the total weight with the unhardened core representing only 3.7 pct of the total weight. Under such circumstances the correction necessary for the more rapid wear rate of this core will probably be less than 1 pct when applied to the average wear rate of the entire ball.

Surface decarburization to a depth of 0.030 in. may increase the rate of wear of a ball 3 in. in diam by a greater amount than a soft core 1 in. in diam. Such a layer represents approximately 6 pct of the total weight of the ball. We estimate that such a zone of decarburization will generally shorten the life of a 3 in. ball by 1 to 2 pct. On smaller balls the influence of decarburization would be still greater.

The data and conclusions presented in this paper are supported by wear tests run over a 7 yr period on over 200 metallurgical classifications of steel and iron balls. A total of 94 wear tests was run to study numerous variables under a wide variety of operating conditions. The entire mass of data collected, if presented in this paper, would increase its length excessively and tend to obscure many of the more important findings which we feel the investigation has brought forth. The data presented in this paper will, therefore, be confined principally to illustrative examples rather than to a complete compilation of all results

Rates of Wear and Limits of Accuracy

Once the surface of a test ball has been suitably prepared, the determination of an accurate abrasion factor is merely a matter of running the test ball along with a standard for a sufficient length of time to wear off an accurately measurable amount of metal. In our tests we attempted to wear off a sufficient weight of metal on each test so that our limit of experimental error in weighing was less than 1 pct. In the 3 ft diam test mill at Golden, Colo., we found that a 24 hr test, using a continuous feed of crushed Climax ore, washed river sand or washed river pea gravel would wear 15 to 30 g from a ball 3 in. in diam. For a group of 5 to 15 balls, this was sufficient to obtain the desired degree of accuracy.

When running tests in which it was necessary to add the abrasive and water in batches instead of continuously we usually ran each test for a period of 6 hr with the abrasive being changed every 2 hr. Two or more of these 6-hr tests were generally run in order to obtain a check on our results.

In the tests in commercial mills the average period of the wear test was from one to two weeks which in most cases wore off 75 to 400 g per ball. This was ample to establish wear factors with a high degree of accuracy. In one case, however, in which we ran a test in a mill grinding cement clinker the rate of wear was so low that only 4.5 g were worn off each 3 in. standard ball after a 222 hr run. While the limit of experimental error on this test was relatively high, a compensating factor was found in the fact that the spread in wear resistance between the good and poor types of balls was so great that there was no doubt about the relative merits of the various types tested.

Table 1 gives the actual rates of wear obtained on our standard balls when run in ten tests representing a rather wide variety of operating conditions. From these values of wear rates per unit of area, the rate at which the balls decrease in diameter and their useful life period have been calculated for each condition. This calculation assumes that mill operating conditions remain constant and that the wear on each ball continues to be in direct proportion to its surface area for its entire life. The life of individual balls in a mill is of particular interest when large scale wear tests are planned on one or more types of balls. General practice on such tests is to start adding the test balls to the mills daily in the quantity needed to maintain the ball charge at a desired level. When the operator is satisfied that the balls formerly used in the mill are substantially all worn out, and the test balls have worn in sufficiently to form an equilibrium ball charge, he then starts to keep an accurate record of the weight of test balls added to the mill. It may be seen from Table 1 that this method of testing can develop into a long, tedious process because of the fact that it requires months and in some cases years to replace the former charge of balls with the test balls.

The rate at which individual balls wear will tend to be faster in mills of large diameter than in mills of small diameter. Theoretically the wear rate of an individual ball in a charge will increase as the 0.6 power of the mill diameter. This general trend for the balls to wear faster in the larger diameter mills is observable from the data in Table 1. An accurate experimental

TABLE 1—*Wear Rates on Standard Martensitic Forged Steel Balls in Various Mills*

Company	Location	Abrasive	Mill Diameter (Feet)	Ball Makeup Diameter (Inches)	Wear per 100 Sq Cm per 24 Hrs (Grams)	Loss in Diameter per 24 Hr (Inches)	Approximate* Life (Days)
Climax Molybdenum Co.	Climax, Colo.	Mo ore	9	100 pct—3 in.	22.7	0.0228	99
Climax Molybdenum Co.	Climax, Colo.	Mo ore	9	50 pct—3 in. 50 pct—2 in.	17.1	0.0172	131 on 3 in. 73 on 2 in.
Climax Molybdenum Co.	Climax, Colo.	Mo ore	6	100 pct—3 in.	13.3	0.0134	168
Homestake Mining Co.	Lead, S. Dak.	Gold ore	5	100 pct—2 in.	9.0	0.0090	139
Phelps-Dodge Corp.	Ajo, Ariz.	Copper ore	6½	100 pct—3 in.	8.6	0.0086	262
Miami Copper Co.	Miami, Ariz.	Copper ore	8	100 pct—3 in.	10.4	0.0105	214
Colo. Portland Cement	Portland, Colo.	Dry cement clinker	8	4 in. and smaller	0.24	0.00024	13,500 on 4 in.
Colo. School of Mines	Golden, Colo.	River sand	3	3 in.	8.3	0.0083	271†
Colo. School of Mines	Golden, Colo.	Feldspar	3	3 in.	3.7	0.0037	608†
Colo. School of Mines	Golden, Colo.	Calcite	3	3 in.	0.31	0.00031	7260†

* The "life" of a ball is calculated here as the time required to wear it from its original diameter down to 0.75 in. diameter.

† This life is on a charge consisting entirely of 3 in. balls. If the charge had been an equilibrium charge containing balls graduated from 3 in. diam down to 0.75 in. then the wear rate would have been less and the life longer due to the greater total surface area of such a charge.

determination of this 0.6 power rule is, however, rather difficult because of the numerous other variables which generally enter the picture when we change from a mill of one diameter to that of another diameter.

FUNDAMENTAL FACTORS GOVERNING BALL WEAR IN A MILL

In the development of a suitable testing technique, one of the first things we had to determine was the basis on which we could compare the wear of balls which varied in diameter or weight. Davis[3] had stated that balls wore in direct proportion to their weight (or cube of their diameter). Ellis[1] in most of his tests based his wear rates on weight loss per unit of surface area, that is, he assumed that balls wore in direct proportion to the square of their diameter. Bond[4] states that the balls in his tests wore as the 2.29 power of their diameter. Since both Davis' and Bond's methods of determining the proportionality of ball wear were of an indirect nature, we felt that further evidence on this subject should be obtained by more direct methods of observation. Groups of balls were, therefore, prepared which were metallurgically and chemically similar, the only variable being their diameter. These groups were run on various tests in a number of mills. It was found that the balls in these groups wore in direct proportion to their surface area, that is, in proportion to the square of their diameter. An exception to this surface area rule was found in the case where a few balls 4 to 5 in. in diam were run in a charge which consisted principally of balls 3 in. in diam and smaller. It was found that under such conditions the abnormally large balls tended to segregate to the outside of the mill charge where they would naturally absorb a greater than average amount of energy and, therefore, wore somewhat faster than was called for by the surface area law.

The mills in which these tests were run have varied from 3 to 9 ft in diam. Mill speeds ranged from 65 to 78 pct of critical while pulp densities and pulp levels varied sufficiently to produce a wide variation in the degree of impact to which the balls were subjected. In spite of these wide variations in operating conditions, we have been unable to find any deviation from this surface area law for ball wear except for the case mentioned where the large balls segregated to the outside of the mill.

Since, in the absence of size segregation, grinding balls in a mill will tend to wear in direct proportion to their surface area, the wear per unit of area will be the same on metallurgically and chemically similar balls even though these balls vary in size. Data from a number of our tests which confirm this observation are given in Table 2. It will be noted that within the limits of experimental error, for any given test, the wear per unit of area on balls of various diameters is practically identical.

Since grinding balls tend to wear in direct proportion to their surface area, it also follows that they will tend to lose diameter at a constant rate. This has been very nicely confirmed by Prentice's[5] investigations. Garms and Stevens[6] show a further confirmation of this in Fig 2 of their paper.

Davis[3] based his conclusion that balls wear in proportion to their weight on the screen analysis of a number of ball charges. Prentice[5] has compiled more recent data on the screen analysis of a number of ball charges. His compilation, and unpublished data from the screen analysis of ball charges in our Climax mills, tend to confirm the surface area theory of ball wear.

TABLE 2—*Wear Rates of Balls of Various Diameters in Several Tests**

a. Annealed, Forged Steel, Plain Carbon Balls in 3-ft Diam Mill at Golden. Results from one 24-hr Test in River Sand and one 24-hr Test in Climax Ore

Nominal Diameter (Inches)	Actual Area (Sq Cm)	Grams Lost per 100 Sq Cm	
		Sand	Climax Ore
3	178.1	11.0	10.1
2	78.6	10.6	9.8

b. Martensitic Forged Steel (Standard) Balls in a 5-ft Diam Mill at Homestake Mining Co. Results from 239-hr Test in Partially Ground Homestake Ore

Nominal Diameter (Inches)	Actual Area (Sq Cm)	Grams Lost per 100 Sq Cm
3	170.0	89.8
3	176.6	89.1
2½	124.8	90.3

c. Soft Pearlitic Forged Steel Balls (277 Brinell) in a 5-ft Diam Mill at Homestake Mining Co. Results from a 239-hr Test in Partially Ground Homestake Ore

3	164.9	110.1
3½	224.5	109.7

d. Soft Pearlitic Forged Steel Balls (277 Brinell) in a 6½-ft Diam Mill at Phelps Dodge Corp., Ajo, Ariz. Results from a 168-hr Test in Ajo Ore

2¾	151.4	94.8
3¼	209.0	95.3

e. Martensitic Forged Steel (Standard) Balls in a 3-ft Diam Mill at Golden. Results from one 24-hr Test in River Sand and one 24-hr Test in Climax Ore

Nominal Diameter (Inches)	Actual Area (Sq Cm)	Grams Lost per 100 Sq Cm	
		Sand	Climax Ore
3	163.0	8.28	8.43
2½	113.9	8.31	8.30

f. Martensitic Forged Steel (Standard) Balls in a 9-ft Diam Mill at Climax, Colo. Results from one 162-hr Test in Climax Ore

Nominal Diameter (Inches)	Actual Area (Sq Cm)	Grams Lost per 100 Sq Cm
3	175.5	116.3
3	170.0	115.8

* The results given represent averages. A total of five to twenty balls of each type was run in each test.

The nature of ball wear in a ball mill has been the object of much discussion ever since Davis[3] published his original paper on this subject. It is suggested that our technique of wear testing can be used to yield further experimental data on this problem. Our tests indicate that the wear in a ball charge tends to be uniformly distributed over the surface so that the wear on any one ball is proportional to its surface area. It would seem to follow that the energy induced by rotation of the mill and the grinding effect are similarly distributed.

The Mechanism of Wear

Our wear tests lead us to believe that wear may be classified in all cases as occurring by two mechanisms. One is by the removal of oxide films or other chemical coatings which form on the freshly exposed metallic surface of the wearing part. When wear occurs by this mechanism, the chemical composition of the metal is the dominating factor. A reduction of wear under these conditions can be most readily accomplished by selecting a metal or alloy which forms a hard and adherent oxide film such as that obtained on high chromium alloys.

The other classification of wear involves the removal of the surface of the part as metallic particles. When wear occurs by this mechanism, we believe the controlling factors governing rate of wear are determined by the distribution and characteristics of the micro-constituents in the metal or alloy. In the case of grinding balls, operating in mills of commercial size, our tests indicate that most of the wear occurs by the removal of metallic particles since the microstructure of all balls tested has been the dominant factor in the determination of their wear resistance. For instance, we have found that balls of the same analysis but of different microstructure will generally show a corresponding difference in wear resistance. On the other hand, balls of widely different alloy content but with practically the same microstructure will show relatively small differences in wear resistance.

Ellis[2] has found, under the conditions existing in his laboratory testing of grinding balls, that wear caused by the removal of oxide films was a dominating factor. His tests were run in small jar mills on balls 1 in. in diam. Under these conditions the addition of chromium to his steel or iron composition was found to be very effective in reducing ball wear. We suspect that when wear in grinding balls occurs primarily by the removal of oxide films, the forces causing abrasion are unusually mild. Where larger balls are used in the larger test mills, or in mills of commercial size, the metallic particles worn from balls during a wet grinding operation can generally be removed from the ground pulp by gravity or magnetic concentration, though a complete separation is difficult. Probably the finer particles are rapidly oxidized, so quantitative determinations of metallic iron in the ground pulp may be misleading. These ground pulps, when allowed to stand, will often generate surprisingly large volumes of hydrogen, indicating the reduction of hydrogen ions by the metallic iron. This hydrogen evolution is the basis of one quantitative method of determining the amount of metallic iron in these pulps.

We believe that the oxide films play only a minor part in determining the wear of grinding balls in most ball mills. In acid or corrosive pulps, or where small balls are used in mills of small diameter, the influence of oxide films on the balls may become an important consideration. The microstructure of the steel or iron balls may also have a definite influence on the formation of the oxide films. Metallographists are all familiar with the fact that pearlite etches more rapidly than martensite when exposed to oxidizing acids. We have indications that a similar condition exists during the formation of oxide films on the freshly abraded surfaces of grinding balls. For instance, on a series of batch tests in the 3 ft diam mill at Golden, using crushed Climax ore as the abrasive, when the mill atmosphere was changed from air to a pure oxygen atmosphere, a group of pearlitic steel balls showed a 39 pct increase in rate of wear while the standard martensitic balls showed only a 19 pct increase in rate of wear. On the other hand, when we attempted to reduce the rate of oxidation of the balls by operating in an air atmosphere with an alkaline pulp, the pearlitic balls showed a 10.7 pct decrease in rate of wear while the martensitic balls showed only a 7.4 pct decrease. The martensitic steel was, therefore, less affected by changes in oxidizing conditions than the pearlitic steel. Since the oxygen in an air atmosphere apparently did have some influence on rate of wear, it is reasonable to expect that at least some reduction in rate of wear by oxidation may be achieved by making the steel martensitic.

The changes in rate of wear which we obtained by the use of an oxygen atmosphere or by making the pulp alkaline are not nearly so great as those obtained by Ellis[2] in his small jar mills. We believe the reason for this is that in our tests the wear by removal of oxide films did not represent nearly as great a proportion of the total as it did in Ellis' small laboratory mills.

DEFINITION OF MICROSTRUCTURAL CLASSIFICATIONS

Since the balls in our wear tests were studied with particular reference to their microstructure, a brief definition of the terms used to describe these microstructures is in order.

During the solidification of a medium or high carbon steel or of hypoeutectic compositions within the white iron range, the first constituent to solidify is austenite. Upon cooling, this austenite may transform to pearlite, bainite, or martensite with the product of transformation depending on the temperature at which the austenite transforms. The austenite and its transformation products are often referred to as the "matrix" of the steel or iron, and are, at times, so designated in this paper. During the cooling of the solidified austenite, pro-eutectoid carbides or ferrite may be rejected. Since these constituents are rejected around the austenite grains, they are referred to as grain boundary carbides or ferrite.

Spheroidized carbides, which occurred in several groups of balls which we studied, were produced by reheating operations after casting or forging. They existed in all cases as very small globular particles finely disseminated throughout the matrix of the steel.

Sulphides were observable and easily identified in several cast steels of high sulphur content. Where they occurred as envelopes or partial envelopes around the original austenite grains, they are classed as grain boundary sulphides. Where they occurred as globules within the original austenite grain boundaries they are classed as globular sulphides.

When ferrous alloys within the cast iron range of compositions solidify, there will form around or adjacent to the original austenite grains: carbides, which we shall designate as primary or massive carbides; graphite, which will be so designated; and steadite, an iron phosphorus eutectic. In one case, ledeburite, which is a eutectic of austenite or its transformation products and primary iron carbide, is mentioned

Pearlite is the lamellar product resulting from transformation of austenite at temperatures from the *Ae*1 temperatures (approximately 1350°F (730°C) for most of the compositions studied) and about 1000°F (540°C). It will be further classified into coarse, medium and fine pearlite depending on the size and spacing of the lamellae. Bainite is the acicular product resulting from transformation of austenite at temperatures which are generally below 900°F (480°C) and above 450°F (230°C). For most of our compositions, the bainite was formed between 800 and 500°F (430–260°C). Substantial amounts of retained austenite generally were found along with the bainite. Martensite refers to the hard acicular product formed below the *Ar''* temperature of the steel. This *Ar''* temperature was within a range of 550 to 350°F (290–180°C) for most of the compositions which we studied. There were, however, a few compositions which contained a total alloy content sufficient to depress the Ar'' temperature to values near or below room temperature so that refrigeration was necessary to obtain appreciable transformation from austenite to martensite.

The similarity in appearance between martensite and low temperature bainite is such that we may, in a number of cases, have confused one with the other. On our more recent investigations we have attempted to distinguish between bainite and martensite by determining the temperatures at which the balls transformed with "Tempilstiks" and a magnet.

Spheroidite as described by Payson[7] has also been studied in a few of these wear tests. This structure was obtained in a few of the normalized high carbon, low alloy steels.

WEAR RESISTANCE OF STEEL OF VARIOUS MICROSTRUCTURES

Microstructure seems to be the dominating factor insofar as the wear resistance of steel grinding balls is concerned. Structure of the matrix appears to be most important. Grain boundary carbides and undissolved spheroidized carbides have a minor, though by no means negligible, effect on wear resistance. The effect of massive carbides is variable and appears to depend to some extent on the character of the matrix. Grain boundary and massive carbides have a pronounced influence on the resistance shown by the balls to spalling and breakage under severe impact. Grain boundary ferrite is harmful to wear resistance.

Carbon content plays such a dominant part in determining the microstructure of a steel or iron that its influence will be discussed in this section on microstructure rather than in a later section on the influence of chemical composition on wear resistance.

The relative wear resistance (abrasion factors) of eleven of the more important types of steel and iron representing certain typical microstructures is listed in Table 3. The abrasion factors obtained under seven different operating conditions are given. The eleven analyses are listed in their approximate order of merit.

Generally, in the conduct of our wear tests, a wide variety of types was included in each test. The complete data from two of our more comprehensive tests are listed in Tables 4 and 5 for illustrative purposes. It is from such data as these that we have taken the selected data for Table 3.

In studying the influence of microstructure on wear resistance, our discussion can be conveniently divided into three parts, one dealing with a matrix structure made up of austenite, martensite, or bainite or a combination of these, and the second with a matrix structure of fine, medium or coarse pearlite. The third part deals with structures containing a matrix of spheroidal carbides in ferrite.

1. *Balls with a Matrix of Austenite, Martensite or Bainite*

A well known form of steel containing austenite is found in the 12 to 14 pct manganese, 1.0 to 1.3 pct carbon Hadfield manganese steel which has been reheated after casting to about 1900°F (1040°C) and water quenched. This steel represents a rather stable form of austenite which can be work hardened from its as-quenched hardness value of about 10 R_C to a maximum of 58 R_C. In our work on grinding balls the highest hardness observed on the work hardened surface of Hadfield manganese steel was 54 R_C. We have never been able to detect the transformation of this austenite to ferro-magnetic products (such as martensite) by work hardening of the surface. This applies to both balls and crusher parts. This finding is supported by the work of Goss[8] who concluded that no such products are formed from the work

TABLE 3—*Abrasion Factors of Typical Microstructures When Tested as 3-in. Grinding Balls in Various Mills*

	Conditions of Test						
Mill Location	Lead, S. D.	Golden, Colo.	Climax, Colo.	Climax, Colo.	Ajo, Ariz.	Golden, Colo.	Portland, Colo.
Mill Diameter (Feet)	5	3	6	9	6½	3	8
Mill Speed (rpm)	27	32	25	20	23.1	32	17.5
Pulp Density (pct Solids)	75	73	77	75	70	70	Dry
Discharge Level of Pulp	High	High	High	Low	High	High	Low
Duration of Test (Hours)	239	48	145	162	168	36	222
Type of Ore	Gold	River Sand	Molybdenum	Molybdenum	Copper	Feldspar	Cement Clinker
Principal Abrasives	Iron-Magnesium-Silicate	Quartz and Feldspar	Quartz and Feldspar	Quartz and Feldspar	Feldspar and Quartz	Albite and Orthoclase	Ca. Aluminate Ca. Silicate

Item No.	Microstructure	Hardness Bhn	Carbon Per Cent	Abrasion Factors						
				Lead, S. D.	Golden, Colo.	Climax, Colo.	Climax, Colo.	Ajo, Ariz.	Golden, Colo.	Portland, Colo.
1	Mart., Aust., primary Carb. (sand cast)	601	3.23	95	101	95		84	72	
2	Bainite or Martensite plus Austenite	512–627	0.90–1.01	96		95	96	90	77	
3	Martensite, some retained Aust. (Std.)	652–683	0.80–0.86	100	100	100	100	100	100	100
4	Mart., Aust., primary Carb. (chill cast)	637	3.27		106	104	Spalled			Broke
5	Fine high carbon Pearlite	364	1.01	104	107	110	110	128	244	
6	Fine eutectoid Pearlite	375	0.71	110	110	115	115	132	264	366
7	Austenite (Hadfield Mn Steel)	207	1.14	120	133	130	124	154	364	
8	Coarse Pearlite	269	0.70	123		134	127	158		
9	Fine Pearl., primary Carb. (chill cast)	475	3.20	128		138	Spalled	163		
10	Coarse Pearl, primary Carb. (chill cast)	444–460	2.70–3.20	139	160	154	Broke	175	247	
11	Coarse Pearl, primary Carb. (sand cast)	475	3.20		178	176				

Abbreviations: Mart. = Martensite.
Carb. = Carbide.
Aust. = Austenite.
Pearl = Pearlite.

hardening of Hadfield manganese steel. This steel has shown relatively poor wear resistance in all our wear tests in which it has been included. This is observable from the abrasion factors for item 7 of Table 3, from item 42 of Table 4, and items 31 or 32 of Table 5. The influence of the carbon content on this type of austenitic steel has not been investigated by us.

One should not conclude from the results for Hadfield manganese steel that all types of austenite will have poor wear resistance when tested as grinding balls. Other types of austenite, which tend to transform quite readily to martensite when work hardened, exist at room temperature. Many of the groups of balls which we have tested contained austenite of this latter type. These groups have in all cases stood at the top of the list insofar as wear resistance is concerned, which leads us to believe that those types of austenite which will transform readily to martensite when cold worked are not harmful to wear resistance of grinding balls. This is demonstrated in Table 6 which lists the results obtained from nine steel compositions containing retained austenite.

The groups of balls listed in Table 6 were selected from Table 5. Tables 3 and 4 list a number of additional results on groups which contained retained austenite. It will be noted that these groups all show excellent wear resistance.

Some of the most wear resistant groups of balls in our tests had relatively low initial hardness because of their large amount of retained austenite. These balls have shown a substantial amount of work hardening on their surface as determined by Rockwell C impressions. For instance the balls in item 8 of Table 5 had a hardness of 40–41 R_C before testing. After a series of wear tests, which concluded with the test at Ajo, the balls in this group had surface hardnesses in the range of 50–56 R_C. Similarly, the surface hardness of the balls in item 10 of Table 5 increased from 53–55 R_C to 61–64 R_C. These surface hardness readings are possibly low since the Rockwell C impressions have probably penetrated into the softer metal below the surface. Balls which did not contain retained austenite showed no appreciable work hardening at the surface, as measured by the Rockwell C test.

To study more fully the influence of retained austenite in the matrix of grinding balls, tests were run on five steel compositions of varying austenite stability. The composition, heat treatment and hardness of these steels, together with the abrasion factors obtained from them on a wear test in the test mill at Golden, are given in Table 7. The groups are listed in order of decreasing austenite stability. Group 1-WQ represents austenitic manganese steel water quenched directly from the mold. It was completely austenitic except for a small amount of primary carbide around the dendritic grain boundaries. The relatively high rate of wear which we obtained from this type of austenite is similar to that obtained from items 31 or 32 of Table 5.

Microscopic observation and magnetic permeability tests indicated that groups 2-WQ and 2-AQ were similar and were completely austenitic except for some primary carbides around the dendritic grain boundaries. Magnetic permeability tests indicated that a slight amount of transformation to martensite occurred in group 2-WQR when it was refrigerated to $-70°F$ ($-57°C$). Small specimens of steels 2-WQ and 2-AQ could be work hardened to a maximum of 44 R_C by hammering. This work hardening brought about some transformation to ferromagnetic products, indicating the formation of martensite.

Tests similar to those run on the three groups of composition 2 were also run on the three groups of composition 3. Groups 3-WQ and 3-AQ appeared to be completely austenitic except for the primary carbides around the grain boundaries. Group 3-WQR showed appreciable transformation

TABLE 4—*Relative Rates of Wear of 3-in. Diameter Grinding Balls in a 6 × 6-ft. Mill at Climax, Colo. (May 1941)*

Item No.	No. of Balls	Heat Treatment	Microstructure	Hardness† (Bhn)	Analysis, Per Cent C	Mn	Cr	Mo	Ni	Cu	Si	S	P	Density, G. per cc	Abrasion Factor
(a) Forged Steel															
1	5	Oil Quench from forge, T. 375°F.	Mart., Bain. (?), Aust.	627	1.01	0.44	1.06	0.21			0.34			7.77	95
2	6	Oil Quench from forge, T. 375°F.	Mart., Bain. (?), Aust.	578	1.03	1.32	1.06	0.22			0.40			7.80	97
3	12	Forged, reheated 1525°F., W.Q., T. 375°	Mart., Spheroidized Carbide	683	1.01	0.44	1.06	0.21			0.34			7.81	97
4	7	Oil Quench from forge, T. 375°F.	Mart., Bain. (?), Aust.	683	0.75	0.45	0.42	0.26			0.65			7.75	98
5	15	Forged, reheated, W.Q. T. 300°F.	Martensite, austenite	652	0.86	0.59		0.29			0.26	0.029	0.011	7.81	100 Std.
6	12	Forged, reheated, O.Q. T. 375°F.	Mart., Bain., Aust.	652	0.75	0.45	0.42	0.26			0.65			7.78	100
7	10	Water quench from forge	Martensite	745	0.70	0.66					0.15			7.84	100
8	14	Forged, reheated, W.Q.	Martensite	668	0.71	0.63					0.22	0.033	0.013	7.82	101
9	15	Forged, reheated, O.Q.	Not observed	534	0.86	0.59		0.29			0.26	0.029	0.011	7.83	102
10	3	Delayed O.Q. from forge, T. 450°F.	Mart., Aust., G.B. Carbide	600	1.03	1.32	1.06	0.22			0.40			7.79	103
11	15	Forged, air cooled	Fine pearlite, G.B. Carbide	387	1.03	1.32	1.06	0.22			0.40			7.81	109
12	15	Forged, air cooled	Fine pearlite, G.B. Carbide	364	1.01	0.44	1.06	0.21			0.34			7.80	110
13	10	Oil quench from forge	Fine pearlite	370	0.75	0.60*					0.20*			7.85	110
14	7	Forged, oil quenched, T. 1050°F.	Tempered martensite	387	1.01	0.44	1.06	0.21			0.34			7.80	113
15	6	Forged, oil quenched, T. 1050°F.	Tempered martensite	402	1.03	1.32	1.06	0.22			0.40			7.78	113
16	15	Forged, air cooled	Fine pearlite	387	0.75	0.45	0.42	0.26			0.65			7.79	114
17	15	Oil quench from forge	Fine pearlite	375	0.71	0.63					0.22	0.033	0.013	7.83	115
18	14	Oil quench from forge	Fine pearlite, a little ferrite	375	0.70	0.66					0.15			7.82	116
19	8	Forged, oil quenched, T. 1050°F.	Tempered martensite	387	0.75	0.45	0.42	0.26			0.65			7.80	124
20	14	Oil quench from forge	Bainite, Martensite	408	0.46	1.74								7.83	124
21	10	Forged, air cooled	Not observed	340	0.52	1.45	0.85				0.19	0.042	0.026	7.83	126
22	15	Forged, air cooled	Pearlite	302	0.86	0.59					0.29	0.029	0.011	7.83	130
23	20	Forged, air cooled	Coarse Pearlite	286	0.70	0.88					0.27			7.80	133
24	15	Forged, air cooled	Coarse Pearlite	258	0.71	0.63					0.22	0.033	0.013	7.83	134
(b) Cast Steel															
25	8	Sand Cast, norm. 1800°F., T. 600°F.	Spheroid. Carb., T. Mart.	555	1.10	0.47	5.45	0.51			0.73	0.02*	0.04*	7.69	98
26	4	Sand Cast, norm. 1800°F., T. 600°F.	Spheroid. Carb., T. Mart.	600	0.85	0.48	5.88	0.47			0.70	0.02*	0.04*	7.63	100
27	6	Chill Cast, air cooled	Martensite, Austenite	512	0.84	1.14	2.61	0.43	1.51		0.47	0.02*	0.04*	7.72	101
28	4	Sand Cast, norm. 1800°F., T. 1000°F.	Spheroid. Carb., T. Mart.	555	0.85	0.48	5.88	0.47			0.70	0.02*	0.04*	7.61	103
29	8	Sand Cast, cooled in sand	Aust., Pearl., G.B. Carb.	375	1.10	0.47	5.45	0.51			0.73	0.02*	0.04*	7.68	104
30	9	Chill Cast, air cooled, T. 600°F.	Tempered Martensite	555	0.73	0.95	1.84	0.42	0.99		0.47	0.02*	0.04*	7.75	104

31	5	Sand Cast, norm. 1650°F., T. 600°F.	Tempered Martensite	601	0.78	0.93	1.63	0.48	0.87		0.44	0.02*	0.04*	7.69	104
32	14	Chill Cast, air cooled	Fine pearlite, G.B. Carbide	468	1.36	0.93	1.68	0.39	0.98		0.55	0.02*	0.04*	7.71	104
33	13	Sand Cast, cooled in sand	Fine Pearlite	352	1.06	1.51	2.63	0.53			0.44	0.02	0.04	7.78	106
34	14	Sand Cast, cooled in sand	Fine Pearlite	341	1.06	2.02	3.05	0.26			0.42	0.05	0.01	7.83	107
35	4	Sand Cast, cooled in sand	Pearlite, G.B. Carbide	351	0.85	0.48	5.88	0.47			0.72	0.02*	0.04*	7.63	108
36	4	Sand Cast, cooled in sand, T. 1000°F.	Pearlite, G.B. Carbide	341	0.85	0.48	5.88	0.47			0.70	0.02*	0.04*	7.62	108
37	14	Sand Cast, cooled in sand	Fine Pearlite, G.B. Carbide	388	1.29	1.03	1.60	0.42	0.99		0.41	0.03	0.06	7.77	109
38	7	Sand Cast, cooled in sand	Fine Pearlite	375	0.91	1.04	1.66	0.43	0.78		0.61	0.02	0.04	7.62	110
39	7	Sand Cast, cooled in sand	Fine Pearlite	364	0.86	1.04	1.66	0.43	0.78		0.61	0.02	0.04	7.67	110
40	7	Sand Cast, cooled in sand	Fine Pearlite	369	0.78	0.93	1.63	0.48	0.87		0.44	0.02*	0.04*	7.74	111
41	6	Sand Cast, ann. 1800°F., norm. 1650°F.	Spherodite	351	0.86	1.04	1.66	0.43	0.78		0.61	0.02*	0.04*	7.64	113
42	8	Sand Cast, reheated 1900°F., W.Q.	Austenite	200	1.14	12.38					0.54	0.02*	0.04*	7.72	130

(c) Chill Cast Iron

43	4	Cooled in air	Mart., Aust., massive Carb.	637	3.27	0.50*	1.50*		4.50*		0.50*			7.74	104
44	15	Cooled in air	Pearlite, massive Carbide	475	3.20*	0.40*	0.90*	0.25*			1.70*			7.51	138
45	10	Cooled in air	Pearlite, massive Carbide	460	2.75	0.75*					0.30*	0.15*	0.15*	7.57	153
46	13	Cooled in sand	Coarse Pearl., massive Carb.	477	3.60*	0.50*					0.55	0.10*	0.30*	7.58	154
47	7	Cooled in air	Pearlite, massive Carbide	444	2.70	0.21	0.02	0.02			0.22	0.16	0.12	7.59	163

(d) Sand Cast White Iron and High Metalloid Steels

48	7	Cooled in sand	Mart., Aust., massive Carb.	600	3.23	0.64	1.89		4.26		0.62	0.11	0.10	7.71	95
49	8	Cooled in sand, T. 400°F.	Mart., Aust., massive Carb.	600	3.23	0.64	1.89		4.26		0.62	0.11	0.10	7.72	96
50	15	Cooled in sand	Pearlite, G.B. Sulphide	364	0.99	0.55	2.69	0.35			0.50	0.19	0.33	7.72	119
51	14	Cooled in sand	Pearl., G.B. Sulph. and Carb.	387	1.24	1.03	2.63	0.28			0.36	0.12	0.31	7.72	122
52	15	Cooled in sand	Pearl., mass. Carb., glob. Sulph.	401	1.53	1.68	2.56	0.54			0.47	0.16	0.32	7.71	127
53	15	Cooled in sand	Pearl., mass. Carb., glob. Sulph.	415	1.53	1.66	2.61	0.26			0.49	0.17	0.33	7.73	130
54	15	Cooled in sand	Pearl., G.B. Sulphide	331	1.02	0.50	0.19	0.57	1.04		0.42	0.17	0.30	7.75	130
55	14	Cooled in sand	Coarse Pearl., globular Sulph.	285	1.02	1.07					0.46	0.19	0.33	7.72	141
56	5	Cooled in sand	Pearl., G.B. Sulph. and Carb.	460	1.54	2.92	1.10				0.51	0.15	0.26	7.72	141
57	14	Cooled in sand	Fine, Pearl., massive Carb.	532	3.00	1.51	1.94		2.75		0.64	0.13	0.29	7.70	151
58	15	Cooled in sand	Pearlite, Steadite, mass. Carb.	477	3.00*	0.65*	0.80*	0.25*			0.90*	0.17*	0.40*	7.62	176
59	14	Cooled in sand	Pearl., Ledeburite, mass. Carb.	578	3.20*	0.70*	2.50*			2.50*	0.80*			7.69	182

* Approximate analysis.

† Hardness refers to the hardness of the metal actually removed by wear.

Abbreviations:
W.Q. = Water Quenched
T. = Tempered
Norm. = Normalized
Bain. = Bainite
Carb. = Carbide
Spheroid. = Spheroidized
O.Q. = Oil Quenched
G.B. = Grain Boundary
Mart. = Martensite
Aust. = Austenite
Sulph. = Sulphide
Mass. = Massive
Glob. = Globular

TABLE 5—*Relative Rates of Wear of 3-in. Diameter Grinding Balls in 6½ × 15-ft. Mill at Phelps Dodge Corp., Ajo, Ariz. (November 1941)*

Item No.	No. of Balls	Heat Treatment	Microstructure	Hardness† (Bhn)	Analysis, Per Cent									Density, G per cc	Abrasion Factor
					C	Mn	Cr	Mo	Ni	Cu	Si	S	P		
			(a) Forged Steel												
1	5	Oil quench from forge, T. 375°F.	Mart., Bain., Aust.	627	1.01	0.44	1.06	0.21			0.34			7.77	90
2	6	Oil quench from forge, T. 375°F.	Mart., Bain., Aust.	578	1.03	1.32	1.06	0.22			0.40			7.80	92
3	15	Forged, reheated, W.Q., T. 300°F.	Mart., Aust.	652	0.80	0.60		0.29			0.26			7.81	100 Std.
4	10	Oil quench from forge	Fine pearlite	370	0.75	0.60*					0.20*			7.85	131
5	14	Oil quench from forge	Fine pearl., a little ferrite	375	0.70	0.66					0.15			7.82	133
6	15	Forged, air cooled (2¾ in.)	Coarse pearlite	286	0.70	0.88					0.27			7.80	157
7	15	Forged, air cooled (3¼ in.)	Coarse pearlite	269	0.70	0.88					0.27			7.82	158
			(b) Cast Steel												
8	10	Chill cast, air cooled	Aust., Bain., trace of pearl.	364	0.90	1.68	2.85	0.41			0.58	0.02*	00.4*	7.75	94
9	5	Chill cast, air cooled	Aust., Bain., Pearl.	452	0.96	1.68	3.43	0.32			0.65	0.03*	0.04*	7.77	94
10	5	Chill cast, air cooled	Bain., Aust.	512	0.88	0.97	1.56	0.33	0.59		0.92	0.03*	0.04*	7.72	96
11	1	Chill cast, W.Q. ½ min. air cooled	Bain., Aust., globular Sulph.	555	1.00	1.21		0.30*			0.35*	0.06*	0.17*	7.76	97
12	1	Chill cast, W.Q. ½ min. air cooled	Bain , Aust., globular Sulph.	555	1.01	1.19		0.30*			0.35*	0.06*	0.17*	7.76	98
13	3	Chill cast, water quenched	Mart., Aust., G.B. Sulph.	627	0.95	0.90				0.30*	0.35*	0.06*	0.17*	7.69	98
14	3	Chill cast, water quenched	Mart., Aust., G.B. Sulph.	682	0.90	0.85				0.30*	0.35*	0.06*	0.17*	7.72	99
15	1	Chill cast, water quenched	Mart., Aust., G.B. Sulph.	601	1.00	1.21		0.30*			0.35*	0.06*	0.17*	7.72	100
16	2	Chill cast, water quenched	Mart., Aust., G.B. Sulph.	640	0.94	1.36		0.30*			0.35*	0.06*	0.17*	7.71	100
17	4	Chill cast, water quenched	Mart., Aust., G.B. Sulph.	682	0.78	0.75		0.30*		0.30*	0.35*	0.06*	0.17*	7.68	100
18	20	Chill cast, air cooled	Bain., Aust., Pearl.	582	0.79	0.92	1.32	0.27			0.47	0.05	0.13	7.62	104
19	8	Sand cast, norm. 1800°F., T. 600°F.	Spheroid. Carb., T. Mart.	555	1.10	0.47	5.45	0.51			0.73	0.02*	0.04*	7.69	107
20	5	Chill cast, air cooled	Pearl., Mart., Aust.	534	1.03	1.36	2.00	0.30*		1.00*	0.35*	0.06*	0.17*	7.75	112
21	5	Chill cast, air cooled	Pearl., Mart., Aust.	535	1.03	1.35		0.30*		2.00*	0.35*	0.06*	0.17*	7.76	112
22	13	Chill cast, air cooled	Fine Pearl., G.B. Carbide	468	1.36	0.93	1.68	0.39	0.98		0.55	0.02*	0.04*	7.71	121
23	1	Chill cast, W.Q. ½ min. air cooled.	Bain., Mart., Aust., G.B. Sulphide	444	0.94	1.36		0.30*			0.35*	0.06*	0.17*	7.77	125
24	8	Sand cast, cooled in sand	Aust., Pearl., G.B. Carbide	375	1.10	0.47	5.45	0.51			0.73	0.02*	0.04*	7.68	126
25	13	Sand cast, cooled in sand	Fine Pearlite	352	1.06	1.51	2.63	0.53			0.44	0.02	0.04	7.78	128

TABLE 5—(*Continued*)

Item No.	No. of Balls	Heat Treatment	Microstructure	Hardness† (Bhn)	Analysis, Per Cent									Density, G per cc	Abrasion Factor
					C	Mn	Cr	Mo	Ni	Cu	Si	S	P		
26	7	Sand cast, cooled in sand......	Fine Pearlite	364	0.86	1.04	1.66	0.43	0.78		0.61	0.02*	0.04*	7.67	132
27	6	Sand cast, ann. 1800°F., norm. 1650°F....................	Spheroidite	351	0.86	1.04	1.66	0.43	0.78		0.61	0.02*	0.04*	7.64	135
28	1	Chill cast, W.Q. ½ min., air cooled....................	Pearl., Bain., G.B. Sulphide	363	0.92	0.92					0.35*	0.06*	0.17*	7.77	137
29	5	Chill cast, air cooled..........	Pearl., Mart., G.B. Carb., and Sulph.	401	1.02	1.16		0.30*		1.00*	0.35*	0.06*	0.17*	7.75	143
30	1	Chill cast, W.Q. ½ min. air cooled....................	Pearl., G.B. Sulphide	363	0.92	0.90				0.30*	0.35*	0.06*	0.17*	7.78	153
31	5	Chill cast, reheated 1900°F., W.Q....................	Austenite	207	1.15	12.50					0.60*	0.02*	0.04*	7.75	154
32	5	Chill cast, air cooled..........	Aust., G.B. Carbides	207	1.15	12.50					0.60*	0.02*	0.04*	7.74	155
(c) Sand and Chill Cast Iron															
33	7	Sand cast, cooled in sand......	Mart., Aust., Massive Carb.	601	3.23	0.64	1.89		4.26		0.62	0.11	0.10	7.71	84
34	8	Sand cast, cooled in sand, T. 400°F....................	Mart., Aust., Massive Carb.	601	3.23	0.64	1.89		4.26		0.62	0.11	0.10	7.72	87
35	15	Chill cast, air cooled..........	Pearl., Massive Carbide	512	3.20*	0.40*	0.90*	0.25*			1.70*			7.51	163
36	10	Chill cast, air cooled..........	Pearlite, massive Carbide	460	2.75	0.75					0.30*	0.16*	0.15*	7.57	168
37	10	Chill cast, air cooled..........	Pearlite, massive Carbide	444	3.00*	0.50*					0.60*	0.15*	0.25*	7.69	170
38	27	Chill cast, air cooled..........	Pearlite, massive Carbide	469	2.75*	0.75*					0.30*	0.16*	0.15*	7.57	173
39	15	Chill cast, air cooled..........	Pearlite, massive Carbide	486	2.75*	0.75*					0.30*	0.16*	0.15*	7.55	181
40	7	Chill cast, air cooled........	Pearlite, massive Carbide	444	2.70	0.21	0.02	0.02			0.22	0.16	0.12	7.59	185

* Approximate analysis.

† Hardness refers to the hardness of the metal actually removed by wear.

Abbreviations:
W.Q. = Water Quenched
G.B. = Grain Boundary
Bain. = Bainite
Pearl. = Pearlite
Carb. = Carbide
T. = Tempered
Mart. = Martensite
Aust. = Austenite
Sulph. = Sulphide
Spheroid. = Spheroidized

TABLE 6—*Relative Rates of Wear of Nine Steel Compositions Containing Retained Austenite*

(168 Hr Wear Test in 6½ × 15-ft Mill at Phelps Dodge Corp., Ajo, Ariz.)

Group No.	Heat Treatment	Microstructure	Hardness (Bhn)	Analysis, Per Cent								Abrasion Factor
				C	Mn	Cr	Mo	Ni	Si	S	P	
		(a) Forged Steel										
1	Oil Quenched from forge, T. 375°F	Martensite, bainite, austenite	627	1.01	0.44	1.06	0.21		0.34			90
2	Oil Quenched from forge, T. 375°F	Martensite, bainite, austenite	578	1.03	1.32	1.06	0.22		0.40			92
3	Forged, reheated, W.Q., T. 300°F	Martensite, austenite	652	0.80	0.60		0.29		0.26			100 Std.
		(b) Cast Steel										
4	Chill cast, air cooled	Austenite, bainite, trace pearlite	364	0.90	1.68	2.85	0.41		0.58	0.02*	0.04*	94
5	Chill cast, air cooled	Austenite, bainite, pearlite	452	0.96	1.68	3.43	0.32		0.65	0.03*	0.04*	94
6	Chill cast, air cooled	Bainite, austenite	512	0.88	0.97	1.56	0.33	0.59	0.92	0.03*	0.04*	96
7	Chill cast, W.Q. ½ min. air cooled	Bainite, austenite, globular sulph.	555	1.00	1.21		0.30*		0.35*	0.06*	0.17*	97
8	Chill cast, water quenched	Martensite, austenite, G.B. sulph.	627	0.95	0.90				0.35*	0.06*	0.17*	98
9	Chill cast, water quenched	Martensite, austenite, G.B. sulph.	682	0.90	0.85				0.35*	0.06*	0.17*	99

* Approximate analysis.

TABLE 7—*Relative Rates of Wear on Five Steel Compositions of Varying Austenite Stability*

(24-hour Wear Test in 3 × 2 ft Mill at Golden Washed Pea Gravel Abrasive, 72.5 Pct Solids)

Group No.	Heat Treatment	Microstructure	Hardness* (Rc)	Analysis, Per Cent						Abrasion Factor
				C	Mn	Cr	Mo	Ni	Si	
1-WQ	Chill cast, water quenched	Austenite, grain boundary carbide	12	1.15	13.00				0.60	154
2-WQ	Chill cast, water quenched	Austenite, grain boundary carbide	20	1.04	1.10	10.76	1.11	3.94	1.10	120
2-WQR	Chill cast, water quenched, refrig. −70°F	Austenite, grain boundary carbide	23	1.04	1.10	10.76	1.11	3.94	1.10	120
2-AQ	Chill cast, air cooled	Austenite, grain boundary carbide	22	1.04	1.10	10.76	1.11	3.94	1.10	121
3-WQ	Chill cast, water quenched	Austenite, grain boundary carbide	23	0.96	1.35	9.48	1.08	1.53	0.90	105
3-WQR	Chill cast, water quenched, refrig. −70°F	Austenite, martensite, G.B. carbide	37	0.96	1.35	9.48	1.08	1.53	0.90	105
3-AQ	Chill cast, air cooled	Austenite, grain boundary carbide	23	0.96	1.35	9.48	1.08	1.53	0.90	104
4-AQ	Chill cast, air cooled	Martensite, bainite (?), austenite	51	0.72	0.70	2.60	0.36	0.32	0.74	96
5-WQ	Forged, reheated, water quenched, T. 300°F	Martensite, austenite	62	0.80	0.60		0.29		0.26	100 Std.

* This represents the hardness of the metal before it was cold worked by the wear test.

to martensite on cooling to −25°F (−32°C) and additional transformation when cooled to −70°F (−57°C). This is also indicated by the higher hardness of group 3-WQR. Groups 3-WQ and 3-AQ could be work hardened to 53 R_C by hammering, and also showed substantial transformation to ferro-magnetic products. It is clear from these tests that the austenite in the groups of composition 3 was less stable than that of composition 2. This was apparently caused by the higher nickel content of composition 2, since there was very little difference in the content of the other austenite stabilizing elements (carbon, manganese, and chromium) in the two compositions.

Group 4-AQ contained fairly substantial amounts of austenite of relatively low alloy content. It is reasonable to expect that this austenite would be rather unstable and, therefore, easily subject to transformation to martensite when work hardened.

Group 5-WQ was our standard martensitic forged steel with which all test groups are compared. It represents mainly martensite in a mildly tempered condition. This mild temper consisted of a brief reheat at approximately 300°F (150°C) which each ball received as a result of its removal from the water quench before its center was completely cooled.

The wearing characteristics of the groups listed in Table 7 improve with decreasing stability of the retained austenite and reach an optimum in group 4-AQ which represents a relatively low alloy steel containing appreciable amounts of retained austenite. The type of martensite found in group 5-WQ causes a slight falling off in wear resistance.

Refrigeration had no effect on the wear resistance of group 3-WQR even though it raised the initial hardness from 23 R_C to 37 R_C. However this steel was subjected to cold work by the forces of impact involved in the wear test which may have raised its hardness at the wearing surface to a value above 50 R_C. A substantial amount of work hardening occurred on the surface of all the austenitic balls though to such a shallow depth that we were unable to measure the surface hardness with any reasonable degree of accuracy.

It has been brought out that retained austenite at the wearing surface of low alloy, high carbon steel grinding balls will transform to martensite by the cold working effect which normally exists in a ball mill grinding operation. This retained austenite can also be transformed to martensite by refrigeration though the one result of such treatment, obtained from group 3-WQR in Table 7, does not indicate that transformation in this manner will appreciably improve the wear resistance of grinding balls.

Another method used for transforming retained austenite was to temper it for a short time at 600°F (315°C). The influence of tempering treatments up to 600°F (315°C) on the wear rates and hardnesses of three martensitic or bainitic steel compositions containing retained austenite is shown in Table 8. Tempering these compositions at 375°F (190°C) brought about a slight increase in hardness but very little, if any, change in wear resistance. By tempering the compositions at 600°F (315°C), however, the austenite in the steel was transformed to tempered martensite or bainite. Little overall change in hardness occurred as a result of this treatment. A marked drop in wear resistance did occur, however, with the average rate of wear for the three compositions almost doubling as a result of the 600°F (315°C) temper.

It should be noted that the wear tests listed in Table 8 were batch runs in a crushed feldspar abrasive. This abrasive was selected because abrasion tests run in feldspar tend to show a wide spread between balls of different wearing quality. This allows the detection of relatively minor differences in the wearing quality of balls. In this series of 6-hr tests the amount worn off the balls was quite small so that the limit of experimental error was about ±5 pct. For this reason we have listed wear rates in Table 8 instead of abrasion factors which require a higher degree of accuracy.

The abrasives and operating conditions in many commercial ball mills are such that the loss in wear resistance caused by the tempering of hardened balls at 600°F (315°C) will be much less than that shown for Table 8. For instance in Table 4, items 25, 26, 30, and 31 represent groups of balls which contained martensite or bainite plus retained austenite before they were tempered. Item 25 in Table 4 is also represented as item 19 in Table 5. These groups do not have as good wear resistance as we would expect from them if they had not been tempered. However, their loss in wear resistance was obviously not very great since they still show relatively good abrasion factors.

Further tests, of a more comprehensive nature, on the influence of tempering hardened steel balls should have practical value. Antia, Fletcher and Cohen[9] have demonstrated that three separate tempering reactions occur between 180°F (80°C) and 675°F (355°C). A study of the wear resistance of the hardened steel before and after each of the tempering reactions should be of value.

The influence of carbon content on steels and irons having a matrix of martensite, bainite or austenite is of interest. This variable has not been fully investigated over the entire range of compositions by our tests. Quite a few martensitic steels within the range of 0.70 to 0.90 pct carbon have, however, been tested and found to be remarkably similar in wearing characteristics. It was for this reason that we chose a martensitic steel within this range of carbon content as our test standard.

When the carbon content of a martensitic steel was dropped to 0.60 pct or lower a definite falling off in wear resistance was noted. Martensitic 0.50 to 0.60 pct carbon low alloy steels when tested by grinding

TABLE 8—*Influence of Tempering on the Wear Rates of 4 Groups of Hardened Steel Balls*

(Averages from Two 6-hr Wear Tests in 3 × 2-ft Mill at Golden. Feldspar Abrasive. 70 pct Solids)

Group No.	Heat Treatment	Microstructure	Hardness (Bhn)	Analysis, Per Cent						Wear Rate Grams per 100 Sq Cm per Test
				C	Mn	Cr	Mo	Ni	Si	
1-a	O.Q. from forge	Martensite, bainite (?), austenite	606	1.01	0.44	1.06	0.21		0.34	0.7
1-b	O.Q. from forge temper 375°F. 1 hr	Martensite, bainite (?), austenite	627	1.01	0.44	1.06	0.21		0.34	0.7
1-c	O.Q. from forge temper 600°F. 1 hr	Bainite (?), Tempered martensite	600	1.01	0.44	1.06	0.21		0.34	1.5
2-a	O.Q. from forge	Martensite, bainite (?), austenite	482	1.03	1.32	1.06	0.22		0.40	0.8
2-b	O.Q. from forge temper 375°F. 1 hr	Martensite, bainite (?), austenite	578	1.03	1.32	1.06	0.22		0.40	0.8
2-c	O.Q. from forge temper 600°F. 1 hr	Bainite (?), Tempered martensite	578	1.03	1.32	1.06	0.22		0.40	1.2
3-a	Chill cast, air cooled	Bainite, austenite	642	0.73	0.95	1.84	0.42	0.99	0.47	0.7
3-b	Chill cast, air cooled, tempered 375°F	Bainite, austenite		0.73	0.95	1.84	0.42	0.99	0.47	0.7
3-c	Chill cast, air cooled, tempered 600°F. 1 hr	Bainite, tempered martensite	555	0.73	0.95	1.84	0.42	0.99	0.47	1.3
4-a	Forged, re-heated, water quenched, T. 300°F	Martensite, austenite	652	0.88	0.50		0.20			0.8 Std.

river sand in the test mill at Golden or by grinding molybdenum ore at Climax have been found to wear from 5 to 15 pct faster than our standard 0.80 pct carbon, low alloy martensitic steel.

When the carbon content of a martensitic steel was raised to about 1.0 pct the wear resistance of the steel was somewhat better than that of our standard. Such steels generally have a very substantial amount of retained austenite in their structure.

High carbon steels containing over 0.90 pct carbon suffer from the disadvantage of brittleness and susceptibility to quench cracking. Recent tests have indicated that reduced cracking of the high carbon steel would be obtained by quenching the balls from a temperature just above the *Acl* to produce a structure of spheroidized carbides in martensite. A limited amount of data which we have on this structure indicates that it has somewhat better wear resistance than martensitic steels of eutectoid composition. This is in line with experience in the ball bearing field and with experience on the wearing properties of rock drill bits for percussion drilling.

It has been found that relatively high alloy contents are necessary for the production of a matrix of martensite, bainite, or austenite in high carbon irons. Chromium and molybdenum, at least up to a certain amount, are contained principally within the massive primary carbides, having little effect on the matrix structure. Nickel is effective on the matrix structure and is, therefore, used to produce "Ni-Hard" white irons with a matrix of martensite plus austenite. Chromium is used in these Ni-Hard irons to suppress the formation of graphite in the structure. Two typical Ni-Hard irons are listed as items 1 and 4 in Table 3 and items 48 and 43, respectively, in Table 4.

2. *Balls with a Pearlitic Matrix*

Generally speaking, balls with a pearlitic matrix will be inferior in wear resistance to those with a matrix of martensite or bainite, plus retained austenite. A very large proportion of the grinding balls used commercially is, however, still of the pearlitic type so the characteristics of such balls have been rather fully investigated. The inferior wearing characteristics of the pearlitic steels and irons are quite evident from a study of the data in Tables 3, 4, and 5. It is also evident, however, that there is a rather wide range in wearing characteristics of various pearlitic steels and irons. Some of the harder, high carbon pearlitic steels are capable of giving a fairly good account of themselves in the harder types of ore. In grinding the softer types of abrasives, such as feldspar or calcite, none of the pearlitic steels or irons showed up well in comparison with the standard martensitic steel balls.

Pearlitic steels of eutectoid carbon content tend to wear better with increasing fineness of the lamellar structure which is paralleled by an increase in the hardness of the steel. This is indicated by a comparison of items 6 and 8 in Table 3.

The influence of carbon content on the wear resistance of six pearlitic forged steels of similar hardness is demonstrated by Table 9. The trend towards improved wear resistance as the carbon content is raised from 0.52 to 1.03 pct is very consistent. Further data on the influence of carbon content on the wear resistance of a series of alloyed pearlitic steels of relatively high hardness are given in Table 10. Again the trend towards improved wear resistance as the carbon content increases, in this case up to 1.19 pct, is observable

The results in Table 11 are shown to illustrate the influence of carbon content up to the cast iron range on balls having a pearlitic matrix. Further data on the performance of pearlitic white iron balls are also given in Tables 3, 4, and 5. It will be noted from these data that the introduction of sufficient carbon into a composition to cause the formation of primary massive carbides in its structure generally causes a definite loss in wear resistance.

Data on the influence of carbon content on the wear resistance of pearlitic balls are presented by Prentice.[5] While Prentice does not report the microstructure of the balls he tested, there are certain groups in his list which, because of their analysis, heat treatment and hardness obviously have a pearlitic matrix. Six of these groups are re-listed in Table 12. The superiority of the steel of eutectoid or slightly hyper-eutectoid carbon content is evident.

The most wear resistant balls of the pearlitic type are made of steel compositions of high hardness and hyper-eutectoid carbon content. A maximum hardness of about 477 Bhn was obtained in a purely pearlitic steel when its carbon content was in a range of 1.0 to 1.40 pct. All of our pearlitic steels within this carbon range contained an envelope or partial envelope of pro-eutectoid carbides around the grain boundaries. Such carbides are very effective in causing rapid nucleation of the austenite, so in this respect they promote the development of a pearlitic matrix. If large amounts of the pro-eutectoid carbides precipitate at the grain boundaries they will lower the wear resistance of the steel. It is apparently desirable to retain most of this pro-eutectoid carbon within the pearlitic matrix. The judicious use of alloying elements coupled with rapid cooling through the critical range has been found quite effective in accomplishing this.

The data in Table 10 indicate that a high carbon, low alloy pearlitic steel is capable of showing fairly good wear resistance when tested in Climax ore. This is confirmed by the result for item 32 in Table 4 as well as by numerous unlisted results. It should be mentioned, however, that when the softer types of ore are ground (such as some of the porphyry copper ores) none of the pearlitic steels will compare very favorably with the

TABLE 9—*Influence of Carbon Content on the Wear Rates of Six Pearlitic Forged Steel Compositions*
(145-hr Wear Test in a 6 × 6-ft Mill at Climax, May 1941)

Group No.	Heat Treatment	Microstructure	Hardness (Bhn)	Analysis, Per Cent						Abrasion Factor
				C	Mn	Cr	Mo	Ni	Si	
1	Air cooled from forge	Fine pearlite, grain boundary ferrite	340	0.52	1.45	0.85			0.19	126
2	Oil quenched from forge	Fine pearlite, a little ferrite	375	0.70	0.66				0.15	116
3	Oil quenched from forge	Fine pearlite	375	0.71	0.63				0.22	115
4	Air cooled from forge	Fine pearlite	387	0.75	0.45	0.42	0.26		0.65	114
5	Air cooled from forge	Fine pearlite, G.B. carbide	364	1.01	0.44	1.06	0.21		0.34	110
6	Air cooled from forge	Fine pearlite, G.B. carbide	387	1.03	1.32	1.06	0.22		0.40	109

TABLE 10—*Influence of Carbon Content on the Wear Rates of Five Pearlitic Cast Steel Compositions (Low Metalloid)*
(150-hr Wear Test in a 9 × 8-ft Marcy Low Discharge Mill at Climax, August 1945)

Group No.	Heat Treatment	Microstructure	Hardness (Bhn)	Analysis, Per Cent						Abrasion Factor
				C	Mn	Cr	Mo	Ni	Si	
1	Chill cast, air cooled	Fine pearlite	430	0.74	0.57	1.24	0.20		0.67	117
2	Chill cast, air cooled	Fine pearlite	418	0.91	0.61	1.37	0.21		0.47	111
3	Chill cast, air cooled	Fine pearlite, G.B. carbide	422	1.02	0.76	1.37	0.20		1.46	111
4	Chill cast, air cooled	Fine pearlite, G.B. carbide	426	1.07	0.62	1.20	0.21			111
5	Chill cast, air cooled	Fine pearlite, G.B. carbide	424	1.19	0.71	1.32	0.22		0.96	106

TABLE 11—*Influence of Carbon Content on the Wear Rates of Seven High Metalloid Cast Steels and Irons*
(145-hr Wear Test on a 6 × 6-ft Mill at Climax, May 1941)

Group No.	Heat Treatment	Microstructure	Hardness (Bhn)	Analysis, Per Cent								Abrasion Factor
				C	Mn	Cr	Mo	Ni	Si	S	P	
1	Sand cast, cooled in sand	Pearlite, Grain boundary sulphide	364	0.99	0.55	2.69	0.35		0.50	0.19	0.33	119
2	Sand cast, cooled in sand	Pearlite, G.B. sulphide and carbide	387	1.24	1.03	2.63	0.28		0.36	0.12	0.31	122
3	Sand cast, cooled in sand	Pearlite, mass. carbide, globular sulphide	401	1.53	1.68	2.56	0.54		0.47	0.16	0.32	127
4	Sand cast, cooled in sand	Pearlite, mass. carbide, globular sulphide	415	1.53	1.66	2.61	0.26		0.49	0.17	0.33	130
5	Sand cast, cooled in sand	Pearlite, G.B. sulphide and carbide	460	1.54	2.92	1.10			0.51	0.15	0.26	141
6	Sand cast, cooled in sand	Fine pearlite, massive carbide	532	3.00	1.51	1.94		2.75	0.64	0.13	0.29	151
7	Sand cast, cooled in sand	Pearlite, steadite, massive carbide	477	3.00	0.65	0.80	0.25		0.90	0.17	0.40	176

TABLE 12—*Influence of Carbon Content on the Wear Rates of Six Pearlitic Steels and Irons, from Prentice*[6]
(Wear Test in a 32 × 17-in. Mill at City Deep Ltd., Johannesburg)

Test No.	Heat Treatment	Probable Microstructure	Hardness (Rc)	Analysis, Per Cent						Life (Days)
				C	Mn	Cr	Si	S	P	
28	Forged, air cooled (?)	Fine pearlite	40	0.90	0.89	0.85	0.20	0.03	0.04	147
35	Rolled, air cooled	Fine pearlite	35	0.78	0.87	1.10		0.05	0.07	120
20	Rolled, air cooled	Fine pearlite	34	0.65	0.59	2.10	0.90	0.03	0.04	119
22	Rolled, air cooled	Pearlite	27	0.53	1.53	0.28	0.11	0.03	0.05	113
7	Rolled, water quenched	Pearlite, grain boundary ferrite	33*	0.53	0.30		0.24	0.05	0.05	87
29	Cast, Sand cooled (?)	Pearlite, Massive carbide	44	3.14	0.50	Tr	0.30	0.16	0.13	83

* This was the hardness near the center of these balls. They were probably harder near their surface.

compositions containing a matrix of martensite or bainite, plus retained austenite.

3. *Balls Containing Spheroidal Carbides*

Spheroidal carbides may be developed in steel compositions by tempering martensite or bainite or, under certain circumstances, by direct transformation from austenite as described by Payson, Hodapp and Leeder.[7] Indications from our tests are that these spheroidal structures are not as wear resistant as lamellar (pearlitic) structures of the same hardness. A comparison of the wear resistance of three compositions which were heat treated to develop lamellar (pearlitic) and spheroidal structures of approximately equal hardness is given in Table 13. The poorer wear resistance of the spheroidal (tempered martensite) structures is evident.

TABLE 13—*Relative Wear Rates of Lamellar and Spheroidal Structures of Approximately Equal Hardness*

(145-hr Wear Test in a 6 × 6-ft Mill at Climax, May 1941)

Analysis, Per Cent					Lamellar Structure (Air Cooled from Forge, Pearlitic)		Spheroidal Structure (Oil Quenched, Tempered 1050°F. Tempered Martensite	
C	Mn	Cr	Mo	Si	Hardness (Bhn)	Abrasion Factor	Hardness (Bhn)	Abrasion Factor
0.75	0.45	0.42	0.26	0.65	387	114	387	124
1.01	0.44	1.06	0.21	0.34	364	110	387	113
1.03	1.32	1.06	0.22	0.40	387	109	402	113

An exception to the theory that spheroidal structures tend to have inferior wear resistance may be found in the case of structures in which the spheroidal carbides exist in a matrix of martensite. (*See* (1) *above*.)

THE INFLUENCE OF GRAIN SIZE

In steel compositions the prior austenitic grain size of the structure seems to have no influence on the wear resistance of grinding balls provided the microstructure within the grains is the same. This has been demonstrated on both cast and forged balls of various types. A rough classification of the steel balls we have studied in order of decreasing grain size follows: 1. Sand cast steel, as cast. 2. Chill cast steel, as cast. 3. Forged steel, as forged. 4. Forged or cast steel, reheated between 1400 and 1600°F (760–870°C) and quenched in air, oil or water.

Numerous tests have been run in which all four types of steel, similar in composition and microstructure, were included. No appreciable difference in the wear resistance of the four types has been found. Under severe conditions of impact, however, the martensitic or bainitic coarse-grained types often failed by spalling or breakage. Pearlitic steels, irrespective of their grain size, have not spalled or broken on any of our tests. For the martensitic steels, however, indications are that a grain refining heat treatment is desirable on balls which must withstand severe conditions of impact or combinations of impact plus low rates of wear such as are encountered in the grinding of limestone or other very soft abrasives.

Grain size is an important factor governing the wear resistance of cast white iron balls. This is particularly true for cast white iron with a pearlitic matrix. Sand cast pearlitic white iron balls have always had poorer wear resistance than chill cast pearlitic iron in our tests. It is believed that the very coarse primary carbides which exist in sand cast white iron are responsible for this more rapid rate of wear. These carbides may be more subject to spalling on a microscopic scale.

The wear of white iron with a matrix of martensite plus retained austenite does not appear to be so greatly influenced by grain size. It will be noted in a comparison of items 1 and 4 of Table 3 that this particular sand cast Ni-Hard white iron actually wore less than the chill cast Ni-Hard white iron.

Where severe impact occurs in a ball mill or where very large balls are used to accomplish crushing as well as grinding, breakage or spalling of all types of white iron may be expected. White iron of the Ni-Hard type seems to be somewhat more resistant to this spalling and breakage than the pearlitic types of white iron.

THE INFLUENCE OF ALLOYING ELEMENTS

Since microstructure is the dominating factor determining the wear resistance of grinding balls, we believe the primary function of alloying elements is to assist in obtaining the type of microstructure desired. A secondary function of alloying elements may be to provide corrosion resistance or to facilitate the development of an abrasion resistant oxide film on the balls.

The influence and usefulness of alloying elements will be considered for the two microstructural types of grinding balls which have, in our estimation, shown the best economic possibilities. These are: 1. Balls with a matrix of martensite plus retained austenite or bainite plus retained austenite. 2. Balls with a pearlitic matrix.

Balls of type 1 are generally superior to even the best of those of type 2 insofar as wear resistance is concerned. Impact conditions in certain ball mills may, however, be too severe for balls of type 1 particularly if they are coarse grained.

To produce the type 1 microstructures in grinding balls, they must be of a composition which possesses sufficient hardenability to halt transformation to pearlite during their final air or liquid quench from the austenitic state. To accomplish this without resorting to too drastic a quench it is necessary to have alloying elements present in practically all the sizes of grinding balls down to about 1 in. diam. Many combinations of alloying elements can be used. However, the selection of an alloy or alloy combination together with the type of quench should be such that it introduces a minimum of harmful effects into the balls. The development of high residual stresses or quench cracks and the presence of embrittling elements such as phosphorus, sulphur or excessively high manganese and silicon should be avoided. Overstabilization of the retained austenite in the structures should also be avoided.

Numerous compositions and methods of heat treatment used to produce the type 1 microstructures in steel or iron balls 3 in. in diam are to be found in Tables 4 to 8 inclusive. It should be noted, however, that the groups listed in these tables were tested because they contributed to the study of certain variables. In most cases these groups do not represent the best composition or heat treatment for commercial production. The commercial manufacture of some of these groups would involve an unnecessarily high alloy cost, while others, because of their composition, grain size or heat treatment, would produce balls subject to quench cracking or other forms of brittleness.

The most economical alloying element per unit of hardenability effect is manganese. We have found that this element can be used to advantage in amounts up to about 0.80 pct in grinding balls with the type 1 microstructures. Larger amounts of manganese than this are generally undesirable since they tend to cause quench cracking or spalling and breakage of the balls in service. In cast steel balls, silicon contents in excess of 0.70 pct have been found undesirable for the same reasons. The cause of this may be that such high silicon contents were generally associated with an over-reduced condition during the finishing period in the melting furnace. For hardenabilities greater than that obtainable from 0.80 pct manganese plus 0.70 pct silicon we believe the combined use of chromium, molybdenum and nickel will give the best results. Copper, in amounts up to 0.50 pct, has also been used to advantage in cast steel balls.

To determine the most efficient chromium, molybdenum and nickel combinations for commercial production of cast or forged steel balls with the type 1 microstructures, we have found that Hostetter's method[10] can be applied to good advantage. While Hostetter's method of calculation was developed to determine the most economical combinations, we have found that such combinations have also worked well from a performance standpoint insofar as wear resistance and freedom from quench cracking or breakage in service are concerned. In the interest of obtaining greater toughness in the hardened balls, certain steel ball producers prefer, however, to deviate from Hostetter's formula by replacing a part of the chromium with a further addition of molybdenum. It is thought that since molybdenum has very little effect in lowering the temperature range in which austenite begins to transform to martensite, it should be less likely to induce quench cracking or high residual stresses in balls which are quenched to below their Ar″ temperature.

Interrupted quenching techniques, which tend to avoid the development of quench cracks or high residual stresses, can be used to advantage in the production of alloyed balls with the type 1 microstructure. They are particularly useful on coarse grained steels such as those in the as-cast or as-forged condition. A technique which we have used with good success on eutectoid carbon, low alloy, chromium molybdenum steel balls has involved a quench in oil, water sprays, or moving air until the surface of each ball was down to between 600 and 700°F (305–370°C). The balls contained sufficient alloy content to suppress any transformation to pearlite so that they were still completely austenitic at these temperatures. The balls were then removed from the quench and allowed to cool slowly to room temperature. The transformation from austenite to bainite or martensite occurred during the slow cooling period. Usually, the balls became quite strongly ferro-magnetic while they were still at temperatures between 700 and 500°F (370–260°C), indicating that the transformation product was mostly bainite.

For steel compositions, it has been pointed out that martensite (or bainite) plus an unstable austenite is easily obtained by the addition of relatively small amounts of alloying elements. Whether the further addition of alloying elements is beneficial or not is of interest. Our tests show little improvement for any large addition of an alloying element, particularly when economic factors are considered. Chromium additions beyond that necessary for full hardening are probably beneficial when conditions are such that the removal of oxide films on the balls represents a substantial proportion of the wear. In most of our tests, however, the effect of chromium in amounts greater than that necessary for full hardening has been small.

The addition of large amounts of manganese or nickel to a full hardening steel analysis appears to be undesirable because of stabilization of the retained austenite. Molybdenum additions beyond that necessary for full hardening have shown no appreciable effect on wear resistance.

The full possibilities for the carbide-forming elements when applied in a structure of martensite plus spheroidized carbides have not been explored. Fairly highly alloyed compositions, such as those used in high-speed steel cutting tools, may show superior wear resistance though the field of application for such costly compositions in grinding balls seems to be decidedly limited.

Alloying elements in the 1.0 to 1.4 pct carbon pearlitic balls serve to increase the fineness of the pearlitic structure and also to increase its carbon content by partially suppressing the rejection of pro-eutectoid carbides around the original austenite grain boundaries. Alloying elements such as chromium and molybdenum may have a secondary beneficial effect on pearlitic structures by the introduction of alloy carbides. Since much work remains to be done, about all that can now be said is that combinations of chromium in a range of 1.0 to 3.0 pct plus molybdenum in a range of 0.20 to 0.50 pct have given good results. Manganese has also been used in amounts up to 2.0 pct though our tests have indicated that its use above 1.0 pct has contributed little to the wear resistance of these hard pearlitic steels. The nickel which was present in many of our pearlitic compositions in amounts up to 1.0 pct apparently had little influence on wear rates.

Balls made from 1.0 to 1.4 pct carbon low alloy steels may be subjected to a fairly wide range of cooling rates from their austenitic state, while still developing a hard pearlitic structure. A fairly wide range in total alloy content is also permissible. Such balls will, therefore, lend themselves well to production under conditions where a simple fool-proof heat treatment, such as an air quench, is required.

Because of the presence in white iron of massive primary carbides which tend to rob the matrix of its carbide forming alloying elements, Hostetter's formulas do not apply. A combination of nickel in a range of 3.0 to 4.5 pct plus chromium in a range of 1.5 to 2.5 pct (the Ni-Hard irons) has been used quite successfully to produce white iron with a matrix of martensite plus retained austenite.

The Influence of the Metalloids

As might be expected, the presence of phosphorus and sulphur above the usual limits is undesirable in steel balls. Both elements cause brittleness in steel and also injure the hot working properties of wrought steels. Table 14 gives a comparison between a number of low metalloid, and similar high metalloid, pearlitic alloy steels. The high metalloid compositions are definitely inferior in wear resistance, even though their structure, hardness, and carbon content are similar. In Table 5, a high metalloid pearlitic steel represented by item 30 compared unfavorably with items 4, 5, 25, and 26, which represent low metalloid pearlites of approximately the same hardness.

The martensitic steels appear to be less affected by metalloids than the pearlitic steels insofar as wear resistance is concerned. In Table 5, the martensitic steels of high metalloid content are represented by items 11 to 17. These groups had good wear resistance, but the balls in these high metalloid groups were very subject to quench cracking and also to breakage in our wear tests.

White iron balls generally have a relatively high sulphur and phosphorus content. Whether any improvement in their wear resistance could be obtained by lowering their metalloid content is not known.

The Influence of Abrasives on Wear Resistance

By changing from one type of abrasive to another, very marked differences in the relative wear resistance of the various types of balls have been observed. While the order of merit of a series will generally remain the same, the spread between the performance of a good and poor type will generally be much greater when grinding soft than when grinding hard abrasives. This is readily observable from the data in Table 3. Other tests which have been run

TABLE 14—*Relative Wear Rates of Pearlitic Steels of High and Low Metalloid Content*

(145-hr Wear Test in a 6 × 6-ft Mill at Climax, May 1941)

Group No.	Heat Treatment	Microstructure	Hardness (Bhn)	Analysis, Per Cent								Abrasion Factor
				C	Mn	Cr	Mo	Ni	Si	S	P	
	(a) Low Metalloid Steels											
1	Sand cast, cooled in sand.......	Fine pearlite	352	1.06	1.51	2.63	0.53		0.44	0.02	0.04	106
2	Sand cast, cooled in sand.......	Fine pearlite	341	1.06	2.02	3.05	0.26		0.42	0.05	0.01	107
3	Sand cast, cooled in sand.......	Fine pearlite, G.B. Carbide	388	1.29	1.03	1.60	0.42	0.99	0.41	0.03	0.06	109
4	Sand cast, cooled in sand.......	Fine pearlite	375	0.91	1.04	1.66	0.43	0.78	0.61	0.02	0.04	110
5	Sand cast, cooled in sand.......	Fine pearlite	369	0.78	0.93	1.63	0.48	0.87	0.44	0.02	0.04	111
	(b) High Metalloid Steels											
6	Sand cast, cooled in sand.......	Pearlite, G.B. Sulphide	364	0.99	0.55	2.69	0.35		0.50	0.19	0.33	119
7	Sand cast, cooled in sand.......	Pearlite, G.B. Sulphide and Carbide	387	1.24	1.03	2.63	0.28		0.36	0.12	0.31	122
8	Sand cast, cooled in sand.......	Pearlite, G.B. Sulphide	331	1.02	0.50	0.19	0.57	1.04	0.42	0.17	0.30	130

in calcite, cement rock, and in commercial ores high in feldspar or talc, tend to confirm this. Ellis' data[1] confirm this observation.

A reversal in order of merit may occur among the pearlitic steels and pearlitic white irons when we change from grinding a hard abrasive to the softer abrasives. For instance, in the grinding of quartz or ores high in quartz content, the pearlitic steels are generally superior to pearlitic white iron. When grinding pure feldspar, however, this condition may be reversed. This is observable from a comparison of items 6 and 10 in Table 3.

Possibly an explanation for the way that the various abrasives influence the wear resistance of grinding balls may be obtained by a consideration of the relative indentation hardness of minerals and steel at various hardness levels. Peters and Knoop[11] list the following indentation hardnesses as obtained by microhardness tests with a diamond pyramid:

Calcite......................	135
Steel, Rc 25................	276
Albite Feldspar.............	490
Steel, Rc 47................	496
Orthoclase feldspar.........	560
Quartz......................	710–790
Steel, Rc 65–67.............	791

The foregoing table indicates that quartz has about the same hardness as the hardest steel, so it should be capable of scratching all types of steel. Feldspar minerals, which have a lower hardness than quartz, will not, according to mineralogical theory, scratch the hard martensitic steels but should scratch steels softer than about 50 R_C. This probably accounts for the fact that, in a feldspar abrasive, the martensitic steels show outstanding superiority, whereas, in grinding quartz, which is capable of scratching all steels, the superiority of the martensitic steels is not so great.

A further confirmation of this scratch theory of wear is provided by the wear rates on balls grinding calcite, which is softer than all the steel balls we have studied. Theoretically, no wear should occur on the balls when grinding calcite. Actually some wear did occur in our tests but the wear rate was only about one tenth of that in feldspar and only about one half that of the balls when they were run in water without any abrasive. The calcite, therefore, acted more as a lubricant than as an abrasive.

Other factors which we have found to affect the spread in wear resistance obtained between a good and a poor type of ball are the coarseness of the abrasive and the pulp density in wet grinding operations. An increase in the coarseness of the abrasive has been found to increase this spread. A decrease in pulp density (more dilution by water) was found to increase this spread on a series of tests in the 3-ft diam. mill at Golden using Climax ore or washed river sand as abrasive. Whether this can be taken as a general rule for all mills or abrasives is not known. When the balls were tested in clear water the rate of wear on all balls dropped to about 10 pct of the rate of wear obtained when Climax ore was used as the abrasive.

PRACTICAL ADAPTABILITY OF THE SHORT TIME WEAR TEST

Our use of the short time wear test described in this paper has been both for the study of fundamental factors influencing the wear of balls, and for an evaluation of the merits of the types of balls which are available in commercial quantities. Resistance to wear, spalling and breakage may be studied by this test. Where possible, it is preferable to run these short time tests under normal service conditions in commercial ball mills. The test has been found very useful as a means for the selection and routine inspection of grinding balls. The method could also be applied with minor modifications to the selection and inspection of grinding rods for rod mills.

In the development of alloys for use in parts other than grinding balls which are subject to abrasive wear, this short time wear test may have possibilities provided the wearing forces on the part in question are similar to those on a grinding ball. In this connection we have studied alloy steels for ball mill liners by testing the desired composition and structure in the form of balls 5 in. in diam. Where comparisons are available the relative wear resistance of a given composition and structure as established by this short time test on large balls has shown good agreement with actual service for the same composition and structure when tested as a liner in the same mill.

A number of correlations between the results of our small scale tests on balls and large scale tests in commercial mills are already available. The results from two commercial mills are given in Table 15.

TABLE 15—*Correlation between Relative Rates of Wear. Determined by Large and Small Scale Wear Tests in Two Commercial Mills*

Type of Ball	Relative Rates of Wear*	
	Large Scale Test	Small Scale Test
(a) Primary Grinding of Crushed Molybdenum Ore at Climax, Colorado		
Martensitic Forged Steel (Standard)	100	100
Fine Pearlitic Forged Steel	107–124	108–120
Sand Cast Cr-Mo White Iron	170–180	176
(b) Primary Grinding of Porphyry Copper Ore at Phelps Dodge Corporation, Ajo, Arizona		
Martensitic Forged Steel	100	100
Chill Cast White Iron	180–200	168–185

* Relative rates of wear on the large scale tests were calculated on the basis of the consumption of balls per ton of ore ground to a given fineness. For the small scale tests the relative rates of wear are calculated on the basis of weight lost per unit of surface area of the individual test balls in the same mill.

The agreement between the two types of testing is quite good. In the case of the test at Ajo, the correlation is better when we make allowances for the fact that the metal worn off each white iron ball during the test was somewhat more wear resistant (because of the finer grain size near the chilled surface) than the remainder of the metal in the ball.

Summary

1. By running marked balls in a ball mill along with a group of standard balls of known quality, it is possible to determine the relative merits of any type of grinding ball within a short time.

2. Barring segregation, grinding balls wear in direct proportion to their surface area and therefore decrease in diameter at a constant rate.

3. A matrix of martensite or low temperature bainite, plus retained austenite, has shown the best wear resistance of all the types studied.

4. Spheroidized carbides enhance the wear resistance of a martensitic matrix.

5. The test results indicate that retained austenite is not an undesirable component of the matrix structure providing it is unstable enough to transform to martensite when cold worked.

6. The wear resistance of steel balls improves with increasing carbon content provided it is not rejected as a grain boundary carbide.

7. The prior austenitic grain size of steel balls does not appear to influence wear resistance. Since balls which have a fine austenitic grain size are less prone to fracture or spall from impact, such grain size is advocated for steel grinding balls.

8. Alloying elements affect the wear resistance of steel grinding balls indirectly through their ability to retard the rates of austenite decomposition at subcritical temperatures. The selection of alloying elements should depend upon their ability to develop the desired matrix structure under the conditions of heat treatment selected by the manufacturer.

9. The metalloids sulphur and phosphorus are undesirable constituents in all types of steel balls because they increase brittleness, impair the hot working properties of the steels, increase the tendency of a steel to quench crack, and decrease wear resistance.

10. The primary massive carbides which are formed during the solidification of pearlitic white cast iron are, in general, detrimental to wear resistance. They are more harmful in sand cast than in chill cast balls.

11. The spread between the performance of a structure of good wear resistance and one of poor wear resistance will vary greatly depending on the mineralogical characteristics of the ore or abrasive in the ball mill and also to some extent on such factors as abrasive size, pulp density and degree of impact. It is therefore desirable to run short time ball tests in the same mills where their use on a large scale is being contemplated.

12. Good correlation is obtained between results of short time tests and the results of large scale tests run over long periods of time.

Acknowledgment

We are indebted to the many firms who cooperated with us in supplying the numerous types of balls used for our wear tests. This cooperation together with that of our operating and testing staff at Climax and Golden, Colo., have made this test program possible. We are also particularly indebted to the management and operating staff of the following mining companies who have cooperated with us by running wear tests in their ball mills:

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Phelps Dodge Corp.	Ajo, Ariz.

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Wet Grinding of Ferrosilicon for Heavy Media

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(New York Meeting, February 1948)

THE ferrosilicon grinding unit at the Central Mill of the Eagle-Picher Mining & Smelting Co. near Picher, Okla., was completed in March 1947. The object of the plant was to grind pigs of ferrosilicon, having a size of approximately 1½ by 4 by 12 in. and an analysis of 85 pct Fe and 15 pct Si, to approximately 100 mesh by crushing and wet grinding in a ball-mill classifier circuit. The minus-100 mesh ferrosilicon had formerly been purchased, and was used as medium in the heavy media process for treatment of lead-zinc ores. From the standpoint of dollars and cents, the cost of minus-100 mesh ferrosilicon was roughly $40.00 per ton more than pigs of the corresponding alloy delivered to the mill. Estimated wet grinding costs were considerably below this figure. The possible saving in cost of make-up medium for the cone plant was particularly attractive since, at that time, the plant losses averaged over four tons of minus-100 mesh ferrosilicon per day.

Aside from the probable saving on the grinding charge, the need for such a unit was twofold. First, periodic fluctuations in the differential between the top and bottom densities of the cone were thought to be detrimental to metallurgy. There was little doubt that an intermediate differential existed at which the heavy media separation was most efficient, representing a compromise between concentrate grade and tailing loss. Furthermore, if a finer, but considerably cleaner, medium could be obtained, there was the possibility that the differential could be decreased without increasing the viscosity, or conversely, the viscosity of the medium decreased without increasing the differential. Preliminary to experimental work aimed at pegging the cone differential and circulating medium viscosity at whatever values were found to be most efficient, it was essential to have control over both the non-magnetic content of the medium and the particle size of the contained ferrosilicon. The latter was possible by means of a grinding unit at the mill.

Second, whereas the ferrosilicon alloys are made by a number of manufacturers, only two, to our knowledge, grind it in any quantity for use in heavy media plants. For all practical purposes, the Central Mill was limited to one source of supply, a particularly vulnerable position. In case the supply was cut off, it was thought that pigs of the 85 pct Fe, 15 pct Si alloy could be obtained elsewhere, providing means of grinding them was available.

There was no published information available covering wet grinding of ferrosilicon. Dry grinding in batch mills and in continuous mills with air classification was being practiced, but the cost of grinding was excessive and the dust condition objectionable.

It appeared that primary and secondary crushing was no particular problem, providing heavy duty equipment was used;

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closed circuit wet grinding in a ball mill also looked attractive if classification of the ball-mill discharge could be accomplished. The problem of classification was difficult because of the high settling rate, the lack of supporting slime and the probability of residual or induced magnetism causing flocculation. To overcome these difficulties, provisions were made to add circulating medium for supporting slime, to provide hydraulic classification, and to demagnetize the crushed product, if any of these were found necessary.

The plant was designed to include a 2-ft Symons standard cone, a 4 by 5-ft grate ball mill and a 24-in. Akins classifier. The estimated capacity was 1000 lb per hour of minus-100 mesh product. The grinding unit was completed in March 1947, and experimental work was begun to solve the classification problem.

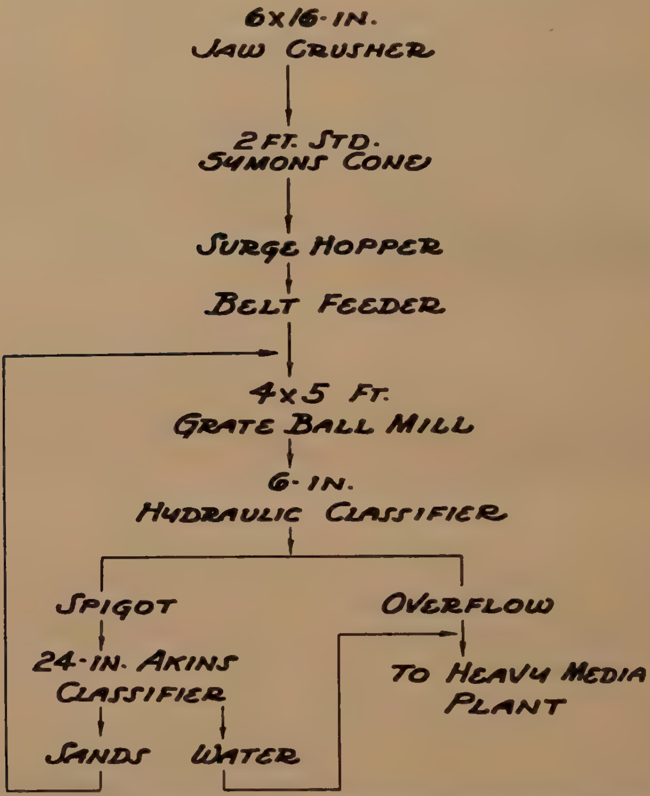

FIG 1—FLOWSHEET OF FERROSILICON GRINDING PLANT, EAGLE-PICHER CENTRAL MILL.

As was expected, the ferrosilicon was practically impossible to classify by density alone. An unstable condition appeared in the classifier pool, characterized by excessive surging. Partial success was attained at high rake speeds by raising the spiral, lowering the weir and adding enough water to the classifier to sluice out the fines. A nonuniform product, representing only about 10 pct of the minus-100 mesh ferrosilicon in the ball-mill discharge, was obtained, and it soon became apparent that either supporting slime would have to be added or a hydraulic classifier constructed. The latter seemed preferable, providing no auxiliary elevating device were necessary.

The vertical distance between the ball-mill discharge and the Akins classifier was 20 in. This required the design of a short hydraulic classifier which would remove the fines from the ball-mill product and discharge the oversize into the Akins classifier. After two weeks of trial and error, a small vortex-type hydraulic classifier was constructed which immediately gave excellent results, and the idea of introducing outside slimes was abandoned. The flowsheet of the plant, as now operated, is shown in Fig 1.

DESCRIPTION OF INDIVIDUAL UNITS

Primary Crushing

The primary breaker is a 16 by 6-in. Blake-type jaw crusher with manganese steel jaw plates and is driven at 308 rpm by a 25 hp motor. The eccentric throw has been reduced to ½ in. with a closed setting of jaws averaging ¾ in. Prior to crushing, the pile of pigs of ferrosilicon is wet down to eliminate dusting and to decrease the tendency of the pigs to jump out of the jaws. The crusher is fed by hand. The conveyor between the jaw crusher and secondary crusher is driven at a speed of 100 fpm by a 2-hp motor and has a slope of 4⅝ in. per foot.

Secondary Crushing

The secondary crusher is a 2-ft specially built, standard Symons cone driven at 355 rpm by a 25-hp motor which consumes 25 to 30 hp under severe load. Bowls and mantles are of manganese steel. The crusher is easily stalled by an overload or by fine dirt contamination from mechanical loading of the stored pigs. The maximum safe feed rate is about 2 tons per hour. The crusher product is uniform and cubical with damp feed, while a dry feed yields material which contains considerable slabs of about twice the desired maximum size. Moisture content is controlled by a water spray mounted above the unit. Closed setting of the crusher averages ¼ in. The conveyor from the Symons to the surge hopper is driven by a 2-hp motor at a belt speed of 60 fpm and is set on a slope of 5⅝ in. per foot. A slight belt slippage occurs with dry material but damp ferrosilicon gives no trouble.

Secondary Storage

The secondary storage bin has a capacity of 60 tons. Since the fine damp material will set up nearly to the hardness of concrete if allowed to remain more than a few days, only a maximum of about 10-tons storage is utilized as a surge ahead of the ball mill. This procedure also decreases difficulty caused by segregation of fines. Minus ¼-in. ball-mill feed occupies about 10 cu ft per short ton.

Fine Grinding

The ball mill is fed by a belt feeder but this might be improved by using a constant weight automatic feeder to smooth out feed irregularities.

The ball mill is a 4 by 5-ft cylindrical, two-trunnion type mill equipped with grate-discharge having ⅜-in. width slots. It has manganese steel, step-type liners measuring 3 ft 8 in. id and a combination 30-in. radius drum and scoop feeder. The ball mill is driven at 32 rpm, 80 pct of critical speed, by a 40-hp slip-ring motor which, under this service, is slightly overloaded. The ball load approximates 40 pct of the volume of the mill. For best results, the ball-mill discharge is carried at about 85 pct solids. Above this figure, the mill tends to load up excessively and stop grinding; below it, the circulating load is excessive and the mill rattles badly. To compensate for ball wear, 8 lb of 2 and 3-in. balls per ton of feed have been fed to the mill. Consumption has not been determined because of the relatively short period of operation, but inspections of the ball charge indicate that the above ball addition is approximately correct.

A screen analysis of the circulating load shows considerable crowding between 65 and 100 mesh, which may be the result of the dearth of smaller balls in the original ball charge. This condition appears to be slowly correcting itself with the addition of about 50 pct of 2-in. balls in the make-up charge, which apparently has had the effect of increasing capacity from the original 1500 lb per hour to 1800 lb per hour of minus-100 mesh material. Capacity of the circuit when grinding to 80 mesh is in excess of 2300 lb per hour. Assorted nails, wires, nuts, bolts, and the like, introduced from unloading box cars, are removed by a ⅛-in. punch plate tramp screen on the ball-mill discharge.

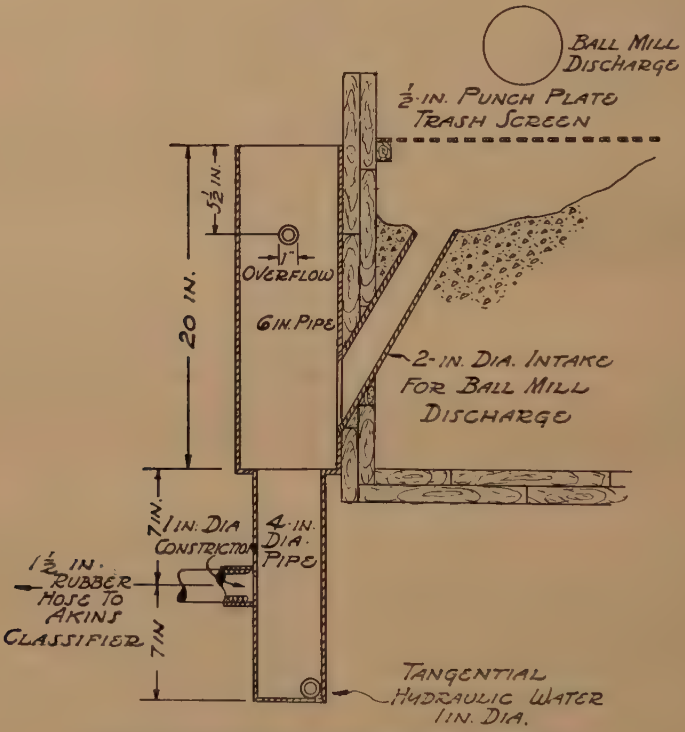
Ball Mill Discharge
½-in. Punch Plate Trash Screen
5½ in.
1"
Overflow
6 in. Pipe
20 in.
2-in. Dia. Intake For Ball Mill Discharge
7 in.
1 in. Dia Constriction
4-in. Dia. Pipe
1½ in. Rubber Hose To Akins Classifier
7 in.
Tangential Hydraulic Water 1 in. Dia.

Fig 2—Construction of Hydraulic Classifier.

Classification

The hydraulic classifier was constructed with the limitation of only 20 in. of head room, and has its hydraulic water introduced tangentially below the spigot. Fig 2 shows the construction of this unit. Most of the water discharges with the sands through the spigot, a 1-in. diam sharp edge constriction, and passes through a 1½-in. hose into the 24-in. Akins classifier. The remaining water passes upward through the sorting column and overflows. The overflow from the hydraulic classifier, ranging from 20 to 30 pct solids, passes down a 1½-in. hose to the overflow box of the Akins. Should an occasion arise when the grain size becomes coarser than desired, the hose is turned back into the Akins pool until the circuit is again in adjustment.

Since the plant is started up daily, this is a useful feature.

The Akins classifier is a 24-in. simplex high-weir type, driven at 6 rpm by a 5-hp motor. A slope of 5⅜-in. per foot is required to gain sufficient head room to drop the sands back into the scoop box. The machine has performed satisfactorily at this angle, and the spiral has not had to be raised at any time for starting, even though the classifier is not drained when shutting down. The launder from the head of the Akins into the scoop box is set on a slope of 10-in. per foot to prevent sanding up.

Particle size in the hydraulic classifier overflow is increased by either adding more hydraulic water or by increasing the circulating load. The cut-off on the maximum particle size with the hydraulic classifier is unusually sharp considering the nature of the feed to it. Removal of the finished product from the ball-mill discharge is also efficiently accomplished.

OPERATING RESULTS

In Tables 1 and 2 are presented screen analyses of the grinding plant products, and also a comparison of the size of the purchased dry-ground ferrosilicon and the wet-ground material produced.

TABLE 1—*Screen Analysis of Grinding Circuit Products*

Tyler Mesh	Opening, Microns	Cumulative Per Cent Retained			
		Ball-mill Feed	Ball-mill Discharge	Akins Sands	Hydraulic Classifier Overflow
½-in.	12,700	0.3			
+ 4	4,760	31.5	0.2	0.5	0
+ 10	1,680	73.4	1.4	4.1	
+ 20	840	83.8	5.5	10.6	
+ 48	297	91.7	35.5	52.2	
+100	149	94.2	68.8	93.3	12.7
+200	74	96.1	80.8	99.0	43.6
+270	53	96.9	85.9	99.5	64.3
+400	37	97.2	90.0	99.6	78.5
−400		2.8	10.0	0.4	21.5

The flexibility of the unit was demonstrated in recent months when it supplied large quantities of coarse (5 pct plus-65 mesh) make-up ferrosilicon to reduce the viscosity of a contaminated medium.

TABLE 2—*Comparison between Various Wet Grinds*

Tyler Mesh	Opening, Microns	Dry Ground	Wet 80-mesh	Wet 100-mesh	Wet 150-mesh
+ 48	297				
+100	149	4.8	12.7	2.8	0.2
+200	74	38.0	30.9	25.0	33.5
+270	53	20.0	20.7	27.0	20.7
+400	37	13.0	14.2	18.0	10.6
−400		24.2	21.5	27.2	35.0

OPERATING DATA

Power Consumption

The total power consumption of the circuit, including crushing, is nearly 100 hp, distributed about half and half between crushing and grinding. The crushing plant operating time is about three-fourths that of the grinding plant. Assuming a capacity of 0.9 tons per hour through 100 mesh, the unit requires roughly 82.0 kw-hr per net ton of minus-100 mesh product produced.

Labor

The plant is operated only one 8-hr shift per day and one man takes care of the entire unit.

Cost

At this early date, the wet grinding cost of ferrosilicon can only be estimated, but apparently will not exceed $5.00 per ton.

SUMMARY

The ferrosilicon grinding unit at the Central Mill has demonstrated the ability to supply a product of suitable size by wet grinding at an indicated cost of 33 pct less than the corresponding dry-ground ferrosilicon purchased in bags or barrels. The wet classification of ferrosilicon in closed circuit operation has been solved with a novel hydraulic classifier. The particle size of its product can be controlled to suit the viscosity of the circulating medium in the heavy media plant, and its value from this standpoint is becoming more apparent daily. Preliminary data on operating conditions are presented.

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